THE BDM CORPORATION

HOLIDAY OFFICE CENTER 3322 MEMORIAL PKWY. S.W. SUITE 32 HUNTSVILLE, ALABAMA 35801 (205) 881-3472

THE BDM CORPORATION

HOLIDAY OFFICE CENTER 3322 MEMORIAL PKWY., S.W., SUITE 32 HUNTSVILLE, ALABAMA 35801 (205) 881-3472

COAL GASIFICATION SYSTEMS ENGINEERING AND ANALYSIS FINAL REPORT APPENDIX C - ALTERNATE PRODUCT FACILITY DESIGNS

December 31, 1980

BDM/H-80-800-TR

This Technical Report is submitted to George C. Marshall Space Flight Center under Contract Number NAS8-33824.

"LEGAL NOTICE"

"This report was prepared by the organization(s) named below as an account sponsored by the Tennessee Valley Authority. Neither TVA, the organization(s) named below, nor any person acting on their behalf: (a) makes any warranty express or implied, with respect to the accuracy, completeness, or usefulness of the information contained in this report, or that the use of any information, apparatus, method, or process disclosed in this report may not infringe privately owned rights; or (b) assumes any liabilities with respect to the use of, method, or process disclosed in this report."

(Organization(s) that prepared this report:)

THE BDM CORPORATION

TABLE OF CONTENTS

APPENDIX C
INTRODUCTION

This appendix contains studies of three different approaches for the production of alternate products from the TVA coal gasification facility. The appendix is divided into three parts. The production of methane, methanol, gasoline, and hydrogen are discussed in three sections as follows:

1. Appendix C-1

   This contains the study of the production of methane, methanol, gasoline, and hydrogen by an add-on facility to a Koppers-Totzek based MBG plant. Applications to a Texaco facility are inferred by evaluation of delta effects from the K-T case.

2. Appendix C-2

   This contains studies of the production of methane from an add-on facility to a Lurgi based MBG plant and of the co-production of methane and methanol from a Lurgi based system.

3. Appendix C-3

   This contains studies of the production of methane from up to 50 percent of the MBG produced in an integrated K-T based plant and the production of methane from up to 50 percent of the MBG produced from an integrated plant in which module 1 is based on K-T technology and modules 2, 3, and 4 are based on Texaco technology.

APPENDIX C-1
PRELIMINARY DESIGNS:
KOPPERS-TOTZEK AND TEXACO
SINGLE PRODUCT FACILITIES

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

LIST OF TABLES (CONTINUED)

LIST OF DRAWINGS

SUMMARY

The purpose of this study was twofold; to provide system level engineering design and evaluation of a Koppers-Totzek and Texaco based gasification facility producing methane, methanol, gasoline, and hydrogen as an alternate to medium Btu gas (MBG), and secondly, to determine the facility capital costs associated with each alternate product to provide relative alternate product costs. A summary of alternate product costs are provided in Table 1 for Koppers-Totzek and Table 2 for Texaco gasification facilities.

Two options were considered for the integration of the add-on alternate-product module(s) with the gasification facility. First, the add-on of the module(s) would be done with minimal integration of the inplant utility and process systems and is the basis of this report. A second, lower level study effort considered a maximum integration of the total facility utility systems and is discussed in another report.

Of the gasification processes considered for the production of MBG from coal, the Koppers-Totzek (K-T) process was dealt with in greatest detail. The various aspects of the Texaco process were evaluated in terms of increments of change from those of the K-T process.

System level costs, based on cost-versus-capacity factoring, are presented on a modular basis. Facility costs were then determined based upon the study design results and single module costs.

Capital costs are provided in 1980 dollars for each system, both onsite and offsite, as well as project contingency, owner's engineering and G&A, and the contractors fee. Operating costs were calculated based on the K-T MBG facility where possible, and on reference literature otherwise.

Detailed technical results are presented for the design of the four add-on modules required for the manufacture of the four alternate products (methane, methanol, gasoline and hydrogen) of interest to TVA. For each process, these results are presented in four major segments;

- Block Flow Diagram
- Process Description
- Material Balance
- Conversion Efficiency.

TABLE 1

ALTERNATE PRODUCTS COST SUMMARY
ADD-ON TO KOPPERS-TOTZEK FACILITY

	MILLIONS OF 1980 DOLLARS			
	METHANE	METHANOL	GASOLINE	HYDROGEN
TOTAL SYSTEM CAPITAL INVESTMENT	1520.1	1508.6	1747.1	1865.3
TOTAL FACILITY INVESTMENT	1853.1	1839.0	2129.7	2273.8
TOTAL DEPRECIABLE INVESTMENT	2344.4	2334.3	2690.2	2858.4
TOTAL CAPITAL REQUIREMENTS	2410.5	2399.8	2759.5	2925.5
ANNUAL OPERATING COSTS	345.0	373.2	379.9	350.3
FACILITY ANNUAL PRODUCTION, 10^{12} BTU	76.11	79.18	61.40	74.97
PRODUCT COST, 1980 $/MILLION BTU	8.03	8.08	11.21	8.94

TABLE 2

ALTERNATE PRODUCT DELTA COST SUMMARY
TEXACO VERSUS K-T BASED FACILITY

	METHANE	METHANOL	GASOLINE	HYDROGEN
TOTAL CAPITAL REQUIREMENT, 10^6 DOLLARS	2827.27	2816.10	3130.31	3547.47
DELTA K-T BASE, %	+ 14.7	+ 14.8	+ 11.8	+ 17.5
TOTAL OPERATING EXPENSE, 10^6 DOLLARS	335.08	364.77	374.24	340.73
DELTA K-T BASE, %	- 3.0	- 2.3	- 1.5	- 2.8
FACILITY ANNUAL PRODUCTION, 10^{12} BTU	83.84	88.65	79.46	94.97
DELTA K-T BASE, %	+ 9.2	+ 10.7	+ 22.7	+21.1
PRODUCT COST, 1980 $/MILLION BTU	7.63	7.54	9.04	7.61
DELTA K-T BASE, %	- 5.0	- 6.7	- 19.4	-14.9

The major tradeoffs considered during this effort include the areas of system selection and system configuration. The systems selected for use in various combinations for the production of the four alternate products include:

- Shift Conversion
- Acid Gas Removal (AGR)
- Methanation
- Methanol Synthesis
- Gasoline Synthesis
- Hydrogen Purification

In the area of system configuration, the most significant tradeoffs were found in the relative position of the compression and gas conditioning systems.

It was found that less than a 5% product cost variation exists among the various tradeoffs studied. Further cost reduction opportunities exist if the ground rules are changed to allow the use of larger scale systems or modules. In addition, at least one clearly viable choice for each alternate product case was found, with the possible exception of gasoline.

The key conclusions obtained from this study can be summarized as follows:

- Methanol, hydrogen, and methane plants that will work can be engineered today.
- Based on many detailed engineering studies, the cost estimates obtained are within normal conceptual design uncertainty limits.
- The add-on system product costs provide an acceptable basis for evaluating market competitiveness of the alternate products.

DESIGN APPROACH

Introduction

The purpose of this study was twofold; to provide system level engineering design and evaluation of a Koppers-Totzek and Texaco based gasification facility producing methane, methanol, gasoline, and hydrogen as an alternate to medium Btu gas (MBG), and secondly, to determine the cost associated with each alternate product to provide relative alternate product costs.

An add-on, modular approach was used as the design basis for design and costing of each alternate product. Each alternate product was based on the output of net MBG, derived from 5,000 Ton/day coal feed to one module of the base facility, as defined by TASK 5.2.1. The facility level costs and capacities were then determined based on a four module facility.

Two options were considered for the integration of the add-on alternate-product module(s) with the gasification facility. First, the add-on of the module(s) would be done with minimal integration of the inplant utility and process systems and is the basis of this report. A second, lower level study effort considered a maximum integration of the total facility utility systems and is discussed in another report.

Of the gasification processes considered for the production of MBG from coal, the Koppers-Totzek (K-T) process was dealt with in greatest detail. The various aspects of the Texaco process were evaluated in terms of increments of change from those of the K-T process. In addition, a qualitative assessment was made of a number of other gasifiers.

In all cases where Task 5.2.1 results were utilized, those results were modified to include the replacement of the MBG-fired boiler with purchased electrical power. This action utilized the results of system level trades and provides a similar basis for comparison with the base cases provided in Task 5.2.1.

System level costs, based on cost-versus-capacity factoring, are presented on a modular basis. Facility costs were determined based upon the study design results and single module costs.

There are eight basic process operations that, used in various combinations, are capable of producing any of the alternate products. These include:

- High Temperature Shift Conversion

- Low Temperature Shift Conversion
- Acid Gas Removal
- Methanation
- Gas Drying
- Methanol Synthesis
- Gasoline Synthesis
- Hydrogen Purification (PSA Adsorption)

Table 3 summarizes the state of development of each of these process operations as it might be applied to each of the alternate products.

The approach to the process design tradeoff study consists of four steps. The first step is to enumerate all of the options that can be considered for the design. The next step is to limit the choices by selecting reasonable options according to explicitly defined criteria. Third, the selected options are further defined through a design process. Finally, the selected options are compared according to explicitly defined criteria and a final choice is made.

To more fully understand this tradeoff study and its level of completeness, it is of interest to know how the available options are identified. The approach taken for each alternate product has five parts. The first is definition of the required process operations (e.g., shift conversion, methanation, etc.). The second is to optimize the operational sequence. Third, the unit operations (e.g., feed preheat, steam injection, catalytic reaction, etc.) for each process operation is identified. Next, the sequence of these unit operations must be defined. Finally, vendor restrictions on and availability of special materials (such as catalysts) must be considered.

KOPPERS-TOTZEK BASED GASIFICATION PLANT

General

Details of the design approach for the four add-on alternate product modules to the K-T gasifier facility are presented in this section. Each of the alternate-product module designs was considered independently. However,

TABLE 3

SYSTEM TECHNOLOGY ASSESSMENT
ADD-ON FACILITIES

SYSTEM IDENTIFICATION	PRODUCT			
	METHANE	METHANOL	GASOLINE	HYDROGEN
90 SHIFT CONVERSION				
HIGH-TEMPERATURE	CP	CP	CP	CP
LOW-TEMPERATURE	-	-	-	CP
91 METHANATION	CA/CP	-	-	-
92 GAS DRYING	CP	CP	CP	CP
93 METHANOL SYNTHESIS	-	CP	CP	-
94 GASOLINE SYNTHESIS	-	-	CA/RD	-
95 HYDROGEN PURIFICATION				
PSA ADSORPTION	-	-	-	CP

KEY: CP = COMMERCIALLY PROVEN, CA = COMMERCIALLY AVAILABLE, RD = READY FOR DEMONSTRATION

the results presented here do not include unnecessary redundancies. For example, design tradeoffs for the Shift Conversion unit and Acid Gas Removal unit that must be considered for more than one alternate product will be presented for the first case only.

Methane

The objective of this design is to define the process and costs related to the production of pipeline-quality (high BTU) gas. The input to the process is MBG consisting primarily of H_2, CO, CO_2, H_2O, H_2S and COS. The output is fully interchangeable high BTU gas (HBG) or methane (CH_4). Methane is produced from MBG by the methanation of CO and CO_2 according to the relations:

$$CO + 3H_2 \rightleftarrows CH_4 + H_2O$$

$$CO_2 + 4H_2 \rightleftarrows CH_4 + 2H_2O$$

The theoretically required H_2 to CO ratio for the methanation process is 3.0.

There are four general design constraints that must be considered in the design of this methane production facility. The first constraint is that the methanation reaction does not go to completion. This effect is minimized if excess H_2 is supplied to the methanation reaction. This implies that the H_2 to CO ratio leaving the shift system must be greater than the 3.0 value previously identified as the theoretically optimum. Hydrogen is produced and the H_2/CO ratio adjusted to 3.2 for this process by means of a shift reaction given by

$$CO + H_2O \rightleftarrows CO_2 + H_2.$$

The second constraint involves gas mixture quality and its interchangeability with pipeline gas. The first factor is that CO content must be less than 0.1 percent by volume. This is achieved due to 99.9 percent of the CO being converted to methane in the methanator. The second factor is the water content of the product gas which must be less than seven pounds of water per million standard cubic feet (SCF) of gas. This is accomplished by use of a triethylene glycol gas drying unit. The final factor is that the higher heating value of the product gas must be greater than 900 BTU/SCF. This is accomplished by a reduction in the amount of CO_2 in the methanation feed stream.

The next constraint is that the methanation catalyst is deactivated by sulfur compounds in the feed gas. Therefore, the amount of H_2S and COS in the methanation feed gas must be less than 0.02 parts per million by volume (ppmv). Use of a deep sulfur removal process provides for this. COS hydrolysis is a possibility to consider in removing COS by conversion to H_2S. However, COS hydrolysis catalysts are not commerically proven.

The final constraint is that the product gas must be delivered at a pipeline pressure of 1,000 psig. Use of a gas compressor will satisfy this requirement.

There were three process-step sequences considered for methane. The first approach was to shift the CO to CO_2, remove the acid gas, and then perform methanation. The advantage here is that proven technology can be used. The disadvantage is that water is condensed from the process stream twice, requiring extra steam.

The second was to combine the shift process with methanation. This results in a process sequence of acid gas removal, combined shift and methanation, and a final CO_2 removal. The advantage of this approach is that the steam injected ahead of the shift stays in the gas. However, an extra CO_2 removal step is required. In addition, this combined step process is not yet commercially demonstrated.

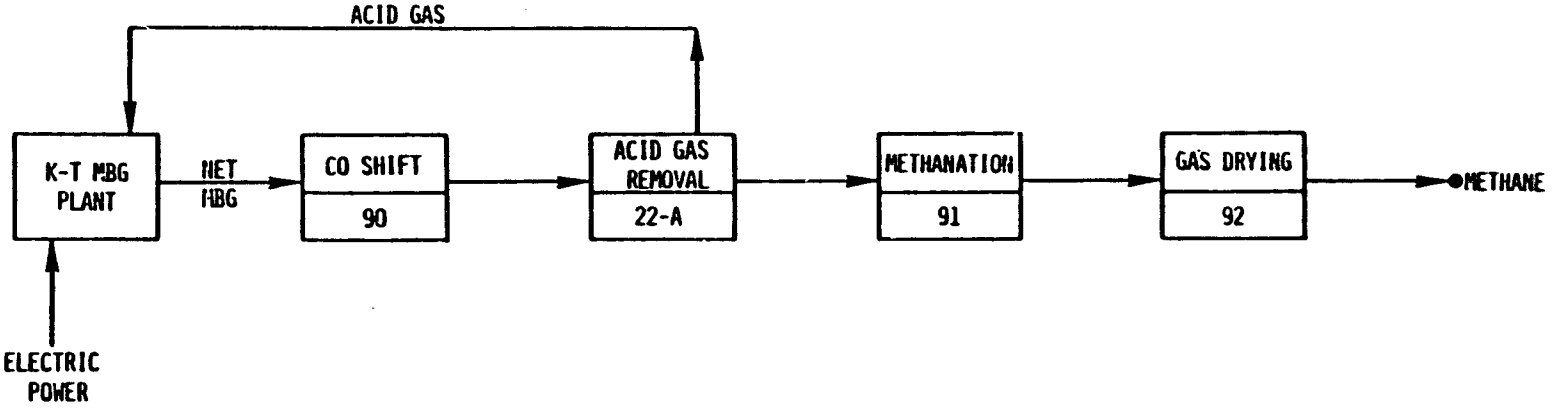
The third approach considered was to modify the first approach such that acid gas removal precedes shift followed by methanation and a final CO_2 removal. The advantage is the same as for the combined shift/methanation scheme discussed above, with the added benefit of being able to use proven equipment. However, there is still an extra CO_2 removal step required.

In evaluating these approaches with respect to each other, a previous team study* was found to be pertinent. This study showed that shift and methanation recovers more high temperature energy than the combined shift/methanation. Also, the extra cost of CO_2 removal makes the third approach of acid gas removal before shift not cost effective. It was, therefore, decided to use the first scheme, as shown in Figure 1.

---

*FE-2240-101

FIGURE 1. KOPPERS-TOTZEK BASE TO METHANE CASE I-A



Once the process steps and their sequence are fixed, it is necessary to more fully define how each process step will be performed. This is done by looking at the various parts of each process step where design choices are available. For the shift conversion process, the design choice is that of the catalyst. Three alternates are available. These are compared to each other by means of a specific set of criteria in Table 4.

An evaluation of this comparison reveals that the iron-chrome catalyst has a lower cost than the cobalt-moly catalyst, while the partial deactivation of iron-chrome offsets its lower cost per pound. Also, the cobalt-moly catalyst offers a possibility of COS hydrolysis, which can lower acid gas removal costs. Finally, a design* done by team members using cobalt-moly is available for this study. It has therefore been decided to use the cobalt-moly catalyst for shift conversion.

There are five acid gas removal systems available. These include Selexol, Rectisol, Benfield, Sulfinol and Stretford. They are compared according to the various criteria in Table 5. A previous team study** shows that the overall attractiveness of the removal processes varies with the partial pressure of the acid gas in the feed stream. In addition, at pressures near those of this study (600 psig), the Benfield, Selexol, and Rectisol processes are preferred. For sulfur removal from a high-sulfur coal gas, selective acid gas removal is preferred. Finally, the Benfield and Selexol processes are limited in their ability to get to less than one ppmv total sulfur with streams containing COS.

Based upon this information, if 200 ppmv sulfur in the treated gas is acceptable, Selexol should be used since Benfield has formate problems. For less than 1 ppmv sulfur, the Rectisol process is required. However, if a proven COS hydrolysis catalyst can be found, then the Selexol process looks slightly more attractive. While the Sulfinol has looked attractive

---

* FE-2240-31

** FE-2240-49

TABLE 4

SHIFT CONVERSION TRADES

ALTERNATIVE NO.	1	2	3
DESCRIPTION	IRON-CHROME	COBALT-MOLY	COPPER
CRITERION	CATALYST (H-T)	CATALYST (H-T)	CATALYST (L-T)
MEETS PERFORMANCE CRITERIA	PARTLY DEACTI-VATED BY SULFUR		NO; CANNOT TOLERATE SULFUR
CAPITAL COST MINIMIZATION	LESS COSTLY/LB	HIGHER SPACE VELOCITY	REQUIRES SULFUR REMOVAL AHEAD
COMMERCIALLY PROVEN	YES	YES	YES
COMPLEXITY	SAME	SAME	SAME
APPLIED DEVELOPMENT NEEDS		TEST COS HYDROLYSIS	
COULD DELAY IMPLEMENTATION	NO	NO	NO
DESIGN DATA AVAILABLE	LITTLE	MOST	LITTLE
OPERATING COST	SAME	SAME	NOT KNOWN
ENVIRONMENTAL PROBLEMS	NO	NO	NO
BYPRODUCT MARKETABILITY	NO	NO	NO

TABLE 5

ACID GAS REMOVAL TRADES

ALTERNATIVE NO.	1	2	3	4	5
CRITERION	SELEXOL	RECTISOL	BENFIELD	SULFINOL	STRETFORD
MEETS PERFORMANCE CRITERIA	VERY COSTLY TO REMOVE COS TO <1PPMV	<1PPMV, COS NO PROBLEM	CANNOT ACHIEVE <1PPMV	MAY ACHIEVE <1PPMV	CANNOT REMOVE COS OR CO_2
CAPITAL COST MINIMIZATION	ABOUT= RECTISOL	VERY COSTLY	INTERMEDIATE COST	INTERMEDIATE COST	HIGH COST
COMMERCIALLY PROVEN	PARTLY	YES	PARTLY	PARTLY	NOT AT PRESSURE
COMPLEXITY	MODERATE	HIGHEST	MODERATE	MODERATE	HIGH WITH COS REMOVAL
APPLIED DEVELOPMENT NEEDS	COS HYDROLYSIS POSSIBLY	NONE	FORMATE PROBLEMS COS HYDROLYSIS	DEEP SULFUR REMOVAL	COS HYDROLYSIS
COULD DELAY IMPLEMENTATION	POSSIBLY	NO	YES	YES	YES
DESIGN DATA AVAILABLE	YES	YES	YES	NO	YES
OPERATING COST	MODERATE	HIGH	MODERATE	MODERATE	HIGH WITH COS REMOVAL
ENVIRONMENTAL PROBLEMS	NO	NO	NO	NO	PURGE LIQUOR
BYPRODUCT MARKETABILITY	-	-	-	-	SULFUR

in past studies, little detailed design information is available at this time. Since deep sulfur removal is required, the decision is to use the Rectisol process. Acid gases were routed to the base plant for processing since the relatively small quantity does not justify a separate sulfur plant.

There are three methanation systems available, including fixed-bed, hot recycle; fixed-bed, cold recycle; and liquid phase. These systems are compared in Table 6. Of the three systems only liquid-phase methanation has not been commercially proven. The hot recycle, fixed-bed scheme uses commercially available catalysts and equipment (except possibly the hot recycle compressor), and recovers more high level energy than the other schemes. A design* done on this methanator by team members is available for use. For these reasons, and since the capital cost differences between hot and cold recycle methanators is not significant, the hot recycle methanator was selected.

Compression and drying is required to meet final pipeline specifications. Drying is a low cost item and needs to be the last process in the overall system. However, the position of the compressor(s) in the system must be determined and several locations are possible. Compression prior to acid gas removal lowers the AGR cost, but requires the compression of the sulfur gases and the CO_2. Compression between the AGR and methanation lowers methanation costs and there is less gas to compress after CO_2 and sulfur gas removal. However, it is still necessary to compress H_2 and CO rather than the CH_4. Compression after methanation requires that only one-half the volume of gas be compressed. Based on these factors, it was decided to compress to pipeline pressure of 1000 psig after methanation. Since the synthesis gas is available at 600 psig.

Methanol

The design objective for methanol production is to produce a fuel-grade methanol with a purity of more than 95% (wt.) from MBG consisting primarily of H_2, CO, CO_2, H_2O, H_2S and COS. Extra high purity (greater than

* FE-2240-31

TABLE 6

METHANATION TRADES

ALTERNATIVE NO.	1	2	3
DESCRIPTION	FIXED-BED	FIXED-BED	LIQUID-PHASE
CRITERION	HOT RECYCLE	COLD RECYCLE	METHANATION
MEETS PERFORMANCE CRITERIA	YES	MAY NEED TRIM METHANATOR	NEEDS TRIM METHANATOR
CAPITAL COST MINIMIZATION	LOWER	LOWER	HIGHER
COMMERCIALLY PROVEN	YES	YES	NO
COMPLEXITY	MODERATE	LEAST	MORE
APPLIED DEVELOPMENT NEEDS	HOT RECYCLE COMPRESSOR, CATALYST	H-T CATALYSTS	TESTING
COULD DELAY IMPLEMENTATION	YES	YES	YES
DESIGN DATA AVAILABLE	DETAILED	BLACK-BOX	BLACK-BOX
OPERATING COST	LOWEST	MODERATE	MODERATE
ENVIRONMENTAL PROBLEMS	-	-	-
BYPRODUCT MARKETABILITY	HP STEAM	HP STEAM	MP STEAM

99% wt.) is not required. Methanol is produced from MBG according to the reaction:

$$2H_2 + CO \rightleftharpoons CH_3OH.$$

The theoretically required H_2/CO ratio for this reaction is 2.0.

There are three general design constraints that were considered in conjunction with this methanol production facility. The first is that the methanol production synthesis reaction does not go to completion. This effect is minimized if excess H_2 is supplied to the synthesis process. This is accomplished by increasing the H_2/CO ratio above the previously indentified value of 2.0 to approximately 2.50. The hydrogen necessary to achieve this ratio is obtained by means of a shift reaction, as it was in the case of methane production. In addition, unreacted gas and inerts in the recycle gas stream lower the conversion efficiency of the reaction. This effect can be minimized by removing CO_2 from the process stream in the AGR system.

A second general constraint is that the methanol synthesis catalyst cannot tolerate sulfur. This requires that total sulfur compounds (H_2S and COS) have a concentration of less than one-half ppmv. This requirement can be met by designing the AGR system for deep sulfur removal.

The final constraint is that higher pressures, on the range of 50 atmospheres, favor reaction equilibrium. This means that compression of the process stream is required.

There are two process step alternatives for a methonol synthesis system. The first is AGR followed by shift, CO_2 removal and methanol synthesis. The second is a shift followed by AGR and then methanol synthesis. An evaluation of these alternatives in a previous team study indicated that the additional cost for another CO_2 removal step in option 1 made option 2 more attractive. Therefore, the second alternative was chosen as shown in Figure 2.

There were three methanol synthesis processes considered. These include low pressure synthesis (50-70 atmospheres), high pressure synthesis (more than 300 atmospheres), and liquid phase methanol synthesis. These are compared with respect to the evaluation criteria in Table 7. An evaluation of these systems indicates that although the per-pass conversion is higher

FIGURE 2. KOPPERS-TOTZEK BASE TO METHANOL CASE I-B.

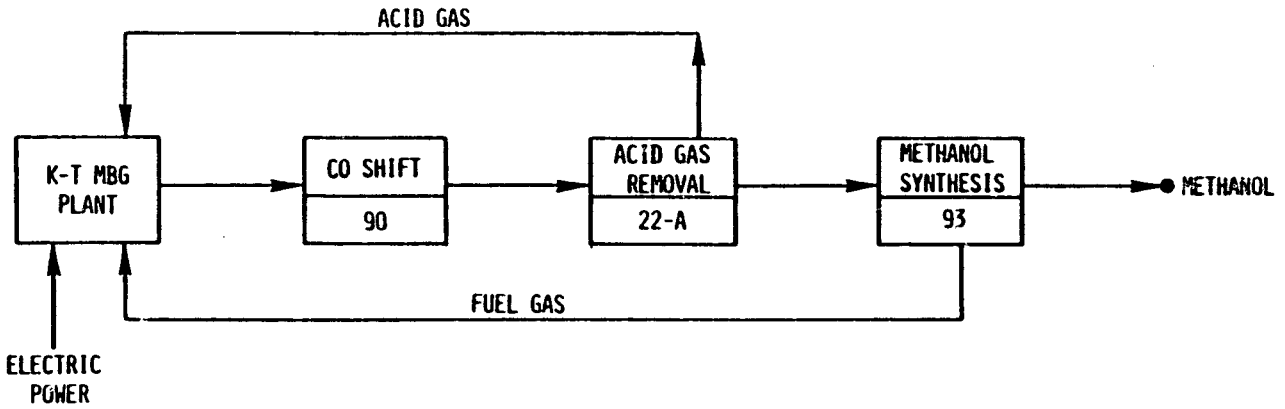


TABLE 7

METHANOL SYNTHESIS TRADES

ALTERNATIVE NO.	1	2	3
CRITERION	LOW-PRESSURE LURGI	LOW-PRESSURE ICI	HIGH-PRESSURE ICI
MEETS PERFORMANCE CRITERIA	YES	YES	YES
CAPITAL COST MINIMIZATION	NO	NO	NO
COMMERCIALLY PROVEN	YES	YES	YES
COMPLEXITY	-	-	-
APPLIED DEVELOPMENT NEEDS	NONE	NONE	NONE
COULD DELAY IMPLEMENTATION	NO	NO	NO
DESIGN DATA AVAILABLE	YES	NO	NO
OPERATING COST	LOWER	LOWER	HIGHEST
ENVIRONMENTAL PROBLEMS	-	-	-
BYPRODUCT MARKETABILITY	-	-	-
COMPARISON AMONG CASES	BEST	-	-

in high pressure synthesis, the syngas compression and recycle energy is over twice that of the low pressure processes. Furthermore the high pressure process is considered obsolete and the liquid-phase process is not yet commercially demonstrated. The Lurgi and ICI low pressure synthesis processes are similar, and design information on the Lurgi process is available to team members. The decision was made to use the Lurgi methanol synthesis process for evaluating the add-on costs of methanol.

Gasoline

The objective of this design is to produce a motor quality gasoline while creating a minimum of byproducts that have maximum marketability.

There are two alternate process-step sequences available. The Mobil M process uses methanol and dehydrates it according to the relationship:

$$xCH_3OH \longrightarrow (CH_2)_x + xH_2O.$$

The second alternate, Fischer-Tropsch synthesis, is direct synthesis according to:

$$xCO + xH_2 \longrightarrow (CH_2)_{x/3} + \frac{x}{3} H_2O + \frac{x}{3} CO_2.$$

Fischer-Tropsch synthesis requires upgrading of the reaction products by means of one or more of the refinery-type processes, including hydrotreating, isomerization, reforming and polymerization.

There are two general constraints associated with gasoline production. First, a motor quality gasoline must be produced and by products minimized. A motor octane of at least 82 and an average of the motor and research octanes of at least 87 must be achieved. Sulfur content must be no more than 0.1 percent by weight. This requires that both processes include an AGR system. The second constraint is that environmental quality must be preserved, which requires that waste waters from both processes be sent through a treatment system.

Both Fischer-Tropsch and the Mobil-M processes consist of three main steps. Each of these steps includes a number of distinct processes. These steps and processes are shown in Table 8.

The two gasoline synthesis processes are compared according to the evaluation criteria in Table 9. The Mobil-M process has been selected over Fischer-Tropsch for a New Zealand synfuels complex. Based on this and a comparative economic study* reviewed by the team, Mobil-M was chosen for the gasoline synthesis process, and is shown schematically in Figure 3.

Hydrogen

The objective of this design is to produce high purity hydrogen (99%) for fuel-cells or hydrotreating service.

The required process steps include shift conversion to produce hydrogen, and acid gas removal to minimize CO_2 content.

There are two general constraints associated with this hydrogen production facility. The first is that the shift conversion process does not go to completion. To minimize the CO content in the shift effluent, the highest activity catalyst is used. The second constraint is that the AGR is not completely effective in removing CO_2 from the process stream. This requires the use of a deep CO_2 removal process.

There are three alternate process-step sequences available. The first is conventional processing which consists of a high temperature shift, followed by sulfur removal, a low temperature shift, CO_2 removal and methanation. A modified conventional process has AGR followed by a low temperature shift, CO_2 removal and methanation. The third alternative, as shown in Figure 4, is AGR followed by a high temperature shift and a pressure-swing-adsorption (PSA) process.

Methanation is used in the first two alternatives to convert the remaining process stream impurities of CO and CO_2 to methane. The PSA unit uses molecular sieves to pass the smaller H_2 molecules through while retaining the larger H_2O, CO and CO_2 molecules. The unit is regenerated by lowering the pressure and purging. This approach results in high losses, 15 to 40%, of the product H_2.

---

*FE-2447-13

TABLE 8

PROCESS SEQUENCE ALTERNATIVES FOR GASOLINE

- FISCHER-TROPSCH
  - GAS PURIFICATION + F-T SYNTHESIS +

GAS PURIFICATION	F-T SYNTHESIS	PRODUCT UPGRADING
--SHIFT	--REACTION	--PRODUCT HYDROTREATING
--AGR	--FRACTIONATION	--PRODUCT FRACTIONATION
	--CATALYST PREPARATION	--C_5/C_6 ISOMERIZATION
		--CATALYTIC REFORMING
		--CATALYTIC POLYMERIZATION
		--POLY GASOLINE HYDROGENATION
		--ALKYLATION
		--ALCOHOL RECOVERY
		--GASOLINE BLENDING

- MOBIL M-GASOLINE PROCESS
  - GAS PURIFICATION + METHANOL SYNTHESIS +

GAS PURIFICATION	METHANOL SYNTHESIS	GASOLINE SYNTHESIS
--SHIFT	--LP OR HP SYNTHESIS (INCL. COMPRESSION)	--SYNTHESIS
--AGR		--FRACTIONATION
		--ALKYLATION
		--WASTEWATER TREATING

TABLE 9

GASOLINE SYNTHESIS TRADES

ALTERNATIVE NO.	1	2
CRITERION	FISCHER-TROPSCH	M-GASOLINE
MEETS PERFORMANCE CRITERIA	YES	YES
CAPITAL COST MINIMIZATION	POORER	BETTER
COMMERCIALLY PROVEN	YES	NO (FIXED-BED READY FOR COMMERCIAL DEMONSTRATION)
COMPLEXITY	HIGHER	LOWER
APPLIED DEVELOPMENT NEEDS	COMMERCIAL U.S. PLANT	COMMERCIAL - SCALE DEMONSTRATION
COULD DELAY IMPLEMENTATION	NO	YES
DESIGN DATA AVAILABLE	YES	YES
OPERATING COST	HIGHER	LOWER
ENVIRONMENTAL PROBLEMS	MORE	LESS
BYPRODUCT MARKETABILITY	POORER	BETTER
COMPARISON AMONG CASES	EQUAL	EQUAL

FIGURE 3. KOPPERS-TOTZEK BASE TO GASOLINE CASE I-C

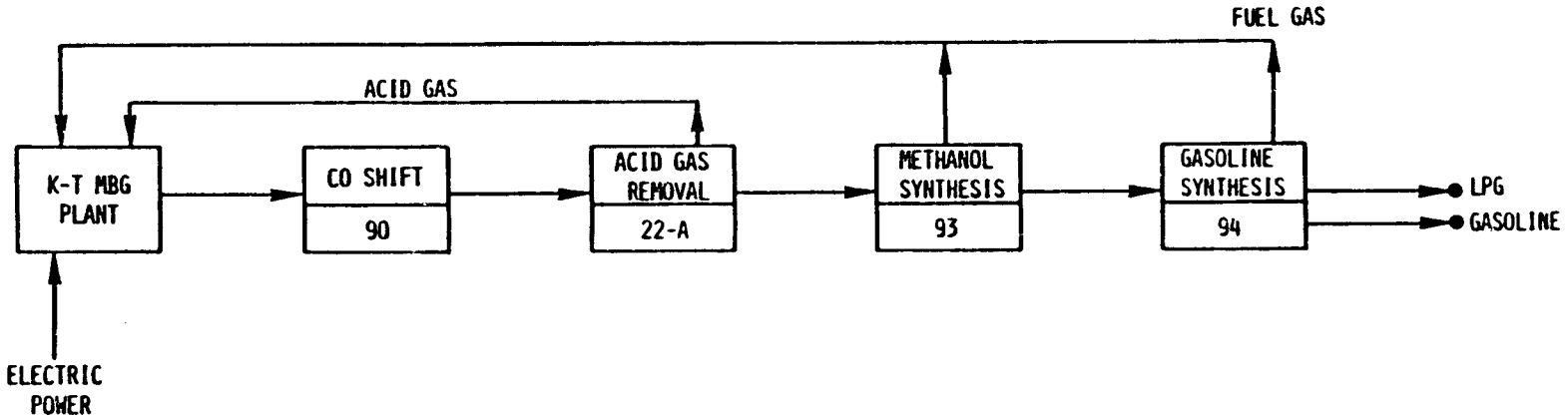


FIGURE 4. KOPPERS-TOTZEK BASE TO HYDROGEN CASE I-D

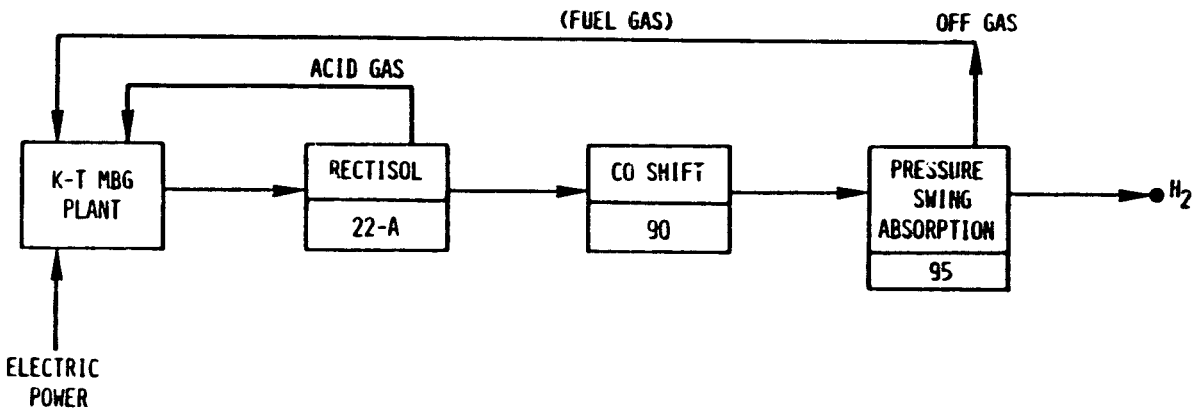


TABLE 10

HYDROGEN PURIFICATION TRADES

ALTERNATIVE NO. CRITERION	1 CONVENTIONAL	2 MODIFIED CONVENTIONAL	3 AGR + SHIFT + PSA
MEETS PERFORMANCE CRITERIA	YES	YES	YES
CAPITAL COST MINIMIZATION	APPROXIMATELY THE SAME		
COMMERCIALLY PROVEN	YES	YES	YES
COMPLEXITY	HIGHEST	HIGHER	LOWER
APPLIED DEVELOPMENT NEEDS	-	-	-
COULD DELAY IMPLEMENTATION	NO	NO	NO
DESIGN DATA AVAILABLE	NO	NO	YES
OPERATING COST	LOWER	LOWER	HIGHER
ENVIRONMENTAL PROBLEMS	-	-	-
BYPRODUCT MARKETABILITY	-	-	-
COMPARISON AMONG CASES	-	-	-

These three sequences are compared in Table 10. An evaluation of the systems includes the fact that the PSA unit has high H_2 losses in the purge gas which reduces net plant output. However, the steam required for CO shift and AGR requires fuel for generation and the purge gas can be used as the required fuel. In addition, design and cost data were not available from the study data base at a level consistent with other systems under study. The decision was made to go with PSA, recognizing the potential adverse impact caused by the H_2 losses.

TEXACO-BASED GASIFICATION PLANT

The technical approach used for the evaluation of the four modules for add-on to the Texaco gasifier facility is discussed here. There are several points to be made in this discussion. First, the basis of the evaluation is the K-T add-on facility costs. Also, as with the K-T case, the Texaco base MBG facility design was adjusted to delete the MBG boilers. Third, the costs of the main process systems in the Texaco add-on facility were scaled from corresponding systems in the K-T case. Next, offsite costs for Texaco were factored from K-T offsite costs based on the ratio of main process costs. Finally, operating costs (other than coal and electrical power) were assumed to be proportional to system capital investment. Coal cost was assumed to be constant and the electrical power cost was obtained by factoring by the ratio of net product BTU's.

EVALUATION RESULTS

In this section, detailed technical results are presented for the design of the four add-on modules required for the manufacture of the four alternate products (methane, methanol, gasoline and hydrogen) of interest to TVA. For each process, these results are presented in four major segments.

The first segment is a block-flow diagram. This diagram not only defines the sequence of processing steps, but also identifies the major process and utility streams within the add-on facility. In addition, a process description is presented for each major process system identified in the block diagram that is required to produce the alternate product from MBG. Third, a material balance is provided in the form of a table. This shows how flow rate and composition of the major streams change throughout the process. It also provides sufficient information to allow proper equipment sizing and cost analysis, as well as calculation of process conversion efficiency.

The final information presented is the expected conversion efficiency (in terms of BTU's out versus BTU's in) for the MBG plant alone, and for the entire plant (MBG plant plus add-on facility). Efficiency data are presented in tabular form. These efficiencies are calculated using coal alone and coal plus electric power as the input to the plant.

Methane

The block flow diagram for methane production from MBG is detailed in Sketch 290-514-SK-001A. The major process units required for the add-on facility include a CO shift, acid gas removal (Rectisol), methanation and gas drying. These units are described in detail below.

The CO shift unit utilizes a high-temperature cobalt-molybdenum catalyst to shift stream composition from carbon monoxide to hydrogen according to the reaction:

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

The inlet gas (MBG product from the K-T plant) has a H_2 to CO ratio of 0.646. 9291 moles of CO and H_2O are reacted to produce a H_2 to CO ratio of 3.20. The shift reaction requires an inlet gas temperature of 660°F and a pressure of 650 psig. The water required for the reaction is added to the process stream in the form of 900°F steam. A steam-to-dry gas mass ratio of 0.65 is required. The process stream is then preheated to the required inlet temperature by means of a heat exchange with the outlet stream.

This shift unit consists of two parallel fixed-bed reactors through which most of the process stream passes. In addition, there are two fixed-bed COS hydrolysis reactors through which flow that portion of the process stream that by-passes the shift catalyst beds. After passing through the CO shift process, the process stream is ready for acid gas removal.

Acid gas removal is accomplished by the Rectisol process licensed by Lurgi. H_2S and COS are removed to less than a one part-per-million by volume (ppmv) concentration. This is accomplished by gas absorption at 600 psig. Sufficient CO_2 is removed so that it constitutes 1% by volume of the effluent gas.

A selective removal design in the Rectisol unit yields one stream of acid gas of at least 50% by volume H_2S, and a waste CO_2 stream suitable for atmospheric venting. Once the sulfur compounds and sufficient CO_2 have been removed, the process stream is passed through a methanation process. The methanator is designed for conversion of 99.9% of the CO and 68.8% of the CO_2 to CH_4. This conversion is accomplished according to the chemical reactions:

$$CO + 3H_2 \rightleftarrows CH_4 + H_2O$$
$$\text{and } CO_2 + 4H_2 \rightleftarrows CH_4 + 2H_2O$$

The methanator consists of two parallel sets of three fixed-bed catalytic reactors in series with intermediate waste heat boilers between them. Temperature control of the reaction is accomplished by means of hot gas recycle. The catalytic reactors are protected from sulfur deactivation by upstream zinc oxide guard beds. The methanator unit also includes compression equipment to pressurize the methanated gas from about 460 to 1000 psig.

The methanated gas next flows to a drying unit. This unit, using triethyleneglycol as a drying agent, removes water from the methanated gas such that the product gas contains no more than seven pounds of water per million standard cubic feet.

The specific effects of the various process units on process stream composition and properties can be obtained from the material balance for methane production from MBG presented in Table 11. Based upon these stream flow rates and composition and the electrical power required by the process, an overall conversion efficiency for the methane product gas is found to be 41.5%. Detailed data are presented in Table 12.

Methanol

The block flow diagram for methanol production from MBG is detailed in Sketch 290-514-SK-001B. The major process units required for this add-on facility include a CO shift, acid gas removal (Rectisol) and methanol synthesis. In addition, there is a tankage requirement for 30 days storage of the methanol product.

The CO shift unit is similar to that required for methanation, except that a H_2 to CO ratio of 2.5 is required for the outlet process stream. Therefore, only 9105 moles of CO need be shifted.

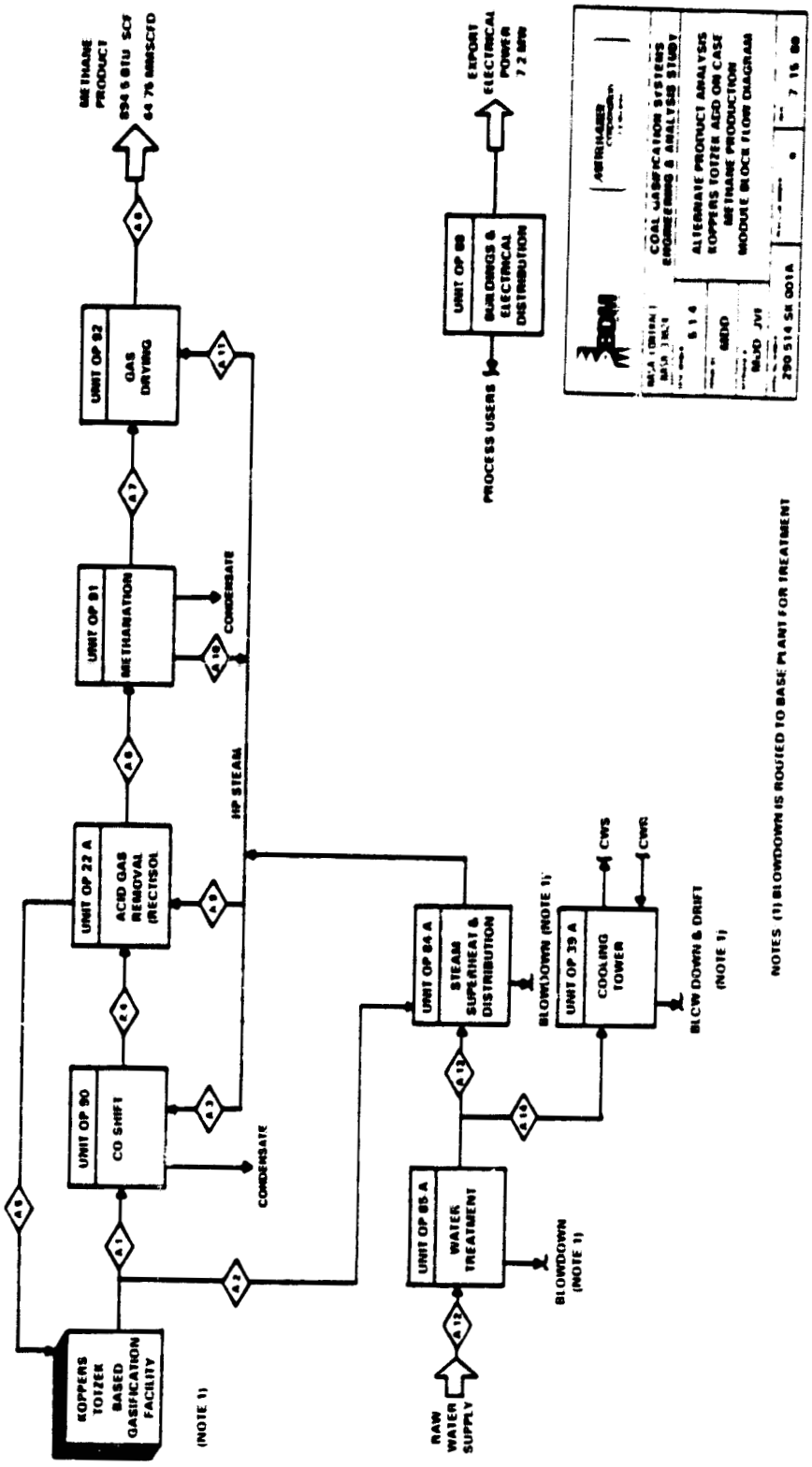


TABLE 11
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE 1-A
KOPPERS-TOTZEK BASE TO METHANE

STREAM NUMBER			A-1	A-2	A-3	A-4	A-5	A-6	A-7	A-8
STREAM NAME / COMPONENT	SYMBOL	M.W.	RAW GAS FEED	FUEL GAS	STEAM TO CO SHIFT	AGR FEED	ACID GAS	METHANATION FEED	GAS DRYING FEED	METHANE PRODUCT
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016	9,865	1,120		19,156	77	19,079	459	459
NITROGEN	N_2	28.016	415	47		415		415	415	415
CARBON MONOXIDE	CO	28.01	15,279	1,735		5,988	11	5,977	6	6
CARBON DIOXIDE	CO_2	44.01	1,452	165		10,743	10,486	257	80	80
HYDROGEN SULFIDE	H_2S	34.076	0.12	0.01		0.12				
METHANOL	CH_3OH	32.042								
CARBONYL SULFIDE	COS	60.075	0.60	0.06		0.60	0.60			
METHANE	CH_4	16.042							6,148	6,148
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_{6}^{+}									
TOTAL DRY	lb MOLES/HR		27,012	3,067		36,303	10,575	25,728	7,108	7,108
WATER	H_2O	18.016	5.5	0.6	15,395	60	60	–	975	1
TOTAL WET	lb MOLES/HR		27,018	3,068	15,395	36,363	10,635	25,728	8,083	7,109
TOTAL	M POUNDS/HR		523.521	59.448	277.386	691.890	461.992	228.816	131.539	114,885
MOLECULAR WEIGHT			19.4	19.4	18.0	19.0	43.4	8.9	16.5	16.2
TEMPERATURE, °F			60	60	900	100	100	75	110	100
PRESSURE, PSIA			615	615	1500	575	575	500	101	1000
MSCFH			10,253	1,164	5,842	13,800	4,036	9,764	3,059	2,698
GPM			–	–	–	–	–	–	–	–

TABLE 11 (Continued)
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE 1-A
KOPPERS-TOTZEK BASE TO METHANE

STREAM NUMBER			A-9	A-10	A-11	A-12	A-13	A-14		
COMPONENT (COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)	SYMBOL	M.W.	STEAM TO AGR	STEAM FROM METHANATION	STEAM TO GAS DRYING	RAW WATER SUPPLY	BOILER FEED WATER MAKEUP	COOLING WATER MAKEUP		
HYDROGEN	H_2	2.016								
NITROGEN	N_2	28.016								
CARBON MONOXIDE	CO	28.01								
CARBON DIOXIDE	CO_2	44.01								
HYDROGEN SULFIDE	H_2S	34.076								
METHANOL	CH_3OH	32.042								
CARBONYL SULFIDE	COS	60.075								
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
HEXANES	C_5H_{10}	70.13								
GASOLINE	C_{6+}									
TOTAL DRY	lb MOLES/HR									
WATER	H_2O	18.016								
TOTAL WET	lb MOLES/HR									
TOTAL	M POUNDS/HR		219.9	725.3	1.7	1200	38.2	1200		
MOLECULAR WEIGHT			18.016	18.016	18.016	18.016	18.016	18.016		
TEMPERATURE, °F			900	596	596	100	250	100		
PRESSURE, PSIA			1500	1560	1500	100	1515	75		
H/S/TH			-	-	-	-	-	-		
GPM			-	-	-	2400	76	2400		

TABLE 12

CONVERSION EFFICIENCY
KOPPERS-TOTZEK TO METHANE

	10^6 BTU/HR INPUTS	10^6 BTU/HR OUTPUTS	EFFICIENCY, PERCENT
COAL-TO-MBG FACILITY	4575		
ELECTRIC POWER TO MBG FACILITY	1345.5		
SUBTOTAL	5920.5		
MBG PRODUCT		3196.2	
COAL TO MBG EFFICIENCY			69.9
OVERALL MBG EFFICIENCY			54.0
MBG TO ALTERNATE PRODUCTS	3196.2		
ELECTRIC POWER TO ALTERNATE PRODUCTS	(98.7)		
ALTERNATE PRODUCT		2413.3	
COAL-TO-ALTERNATE PRODUCT EFFICIENCY			52.8
OVERALL FACILITY EFFICIENCY			41.5

The Rectisol process is used for acid gas removal. Its requirements are the same as those for methane production, except that H_2S and COS must be removed to less than a one-half ppmv concentration.

The methanol synthesis unit is based on a low pressure process licensed by Lurgi. This unit compresses the inlet process stream to 1100 psig. The stream then passes through two parallel fixed-bed reactors where methanol is produced according to the reactions:

$$CO + 2H_2 \rightleftharpoons CH_3OH$$
$$CO_2 + 3H_2 \rightleftharpoons CH_3OH + H_2O$$

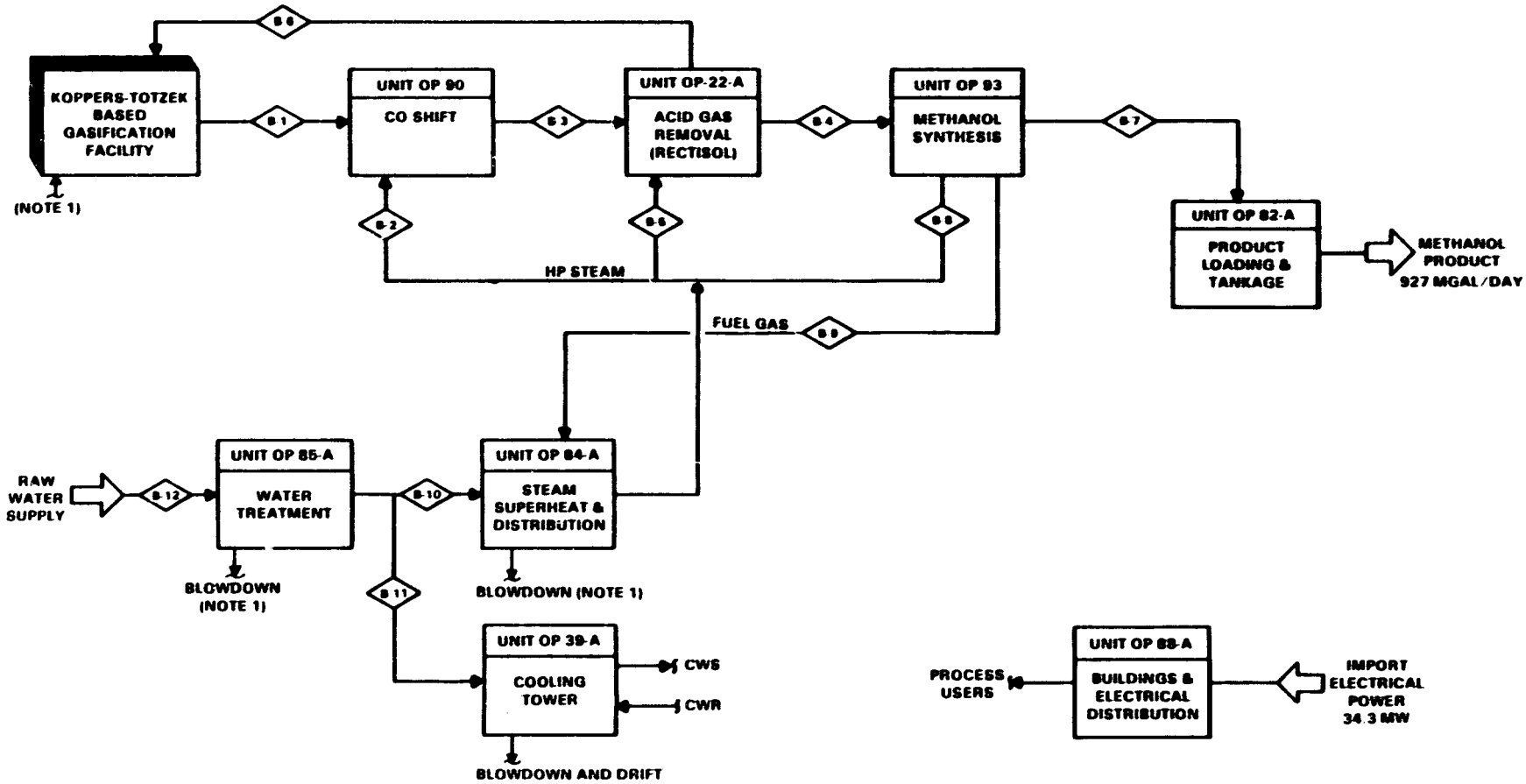
Following the reaction, waste heat is recovered and a portion of the outlet stream is compressed and recycled. The final product stream is 95% methanol by weight and is produced at a rate of 3143 short tons per day.

The material balance for methanol production from MBG is presented in Table 13. The conversion efficiency for the methanol production process is 39.3%. Detailed data are presented in Table 14.

Gasoline

The block flow diagram for gasoline production from MBG is presented in Sketch 290-514-SK-001C. The major process units required for this add-on facility include all those required for methanol synthesis (CO shift, acid gas removal and methanol synthesis) plus gasoline synthesis and biological treatment. In addition, there is a tankage requirement for 30 days storage of the gasoline and byproduct liquids. It should be noted here that this process yields not only a gasoline product, but also liquid petroleum gases (LPG).

As might be expected, since methanol is an intermediate product of gasoline production, the CO shift, acid gas removal and methanol synthesis units are identical to those described above for the production of methanol.



NOTES: (1) BLOWDOWN IS ROUTED TO BASE PLANT FOR TREATMENT

NASA CONTRACT NAS8-33824	COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.1.4	ALTERNATE PRODUCT ANALYSIS KOPPERS-TOTZEK ADD-ON CASE METHANOL PRODUCTION MODULE BLOCK FLOW DIAGRAM
MDD/	
MDD/JVF	
290 514 SK-0018	0 — 7/15/80

TABLE 13
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE 1-B
KOPPERS-TOTZEK BASE TO METHANOL

STREAM NUMBER			B-1	B-2	B-3	B-4	B-5	B-6	B-7	B-8
COMPONENT / STREAM NAME	SYMBOL	M.W.	RAW GAS FEED	STEAM TO CO SHIFT	AGR FEED	METHANOL SYNTHESIS FEED	STEAM TO AGR	ACID GAS	METHANOL PRODUCT	STEAM FROM METHANOL SYNTHESIS
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016	10,985		20,090	20,082		8	-	
NITROGEN	N_2	28.016	462		462	462		-	-	
CARBON MONOXIDE	CO	28.01	17,014		7,909	7,894		15	-	
CARBON DIOXIDE	CO_2	44.01	1,617		10,722	1,050		9,672	17	
HYDROGEN SULFIDE	H_2S	34.076	0.13		-	-		0.13	-	
METHANOL	CH_3OH	32.042	-		-	-		-	7,776	
CARBONYL SULFIDE	COS	60.075	0.60		0.60			0.60		
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR		30,079		39,184	29,488		9,696	7,793	
WATER	H_2O	18.016	6.1	15,087	65			65	501	
TOTAL WET	lb MOLES/HR		30,085	15,087	39,249	29,488		9,761	8,294	
TOTAL	M POUNDS/HR		582.966	271.807	748.063	320.750	271.8	461.141	258.933	341.6
MOLECULAR WEIGHT			19.4	18.0	19.1	10.9	18.016	43.7	31.2	18.016
TEMPERATURE, °F			60	900	100	75	900	100	95	596
PRESSURE, PSIA			615	1500	575	500	1500	575	20	1515
DESCRIPTION			11.417	-	14.895	11.191	-	3.704	3.148	-
GPM							-			-

TABLE 13 (Continued)
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE I-B
KOPPERS-TOTZEK BASE TO METHANOL

STREAM NUMBER			B-9	B-10	B-11	B-12				
STREAM NAME / COMPONENT	SYMBOL	M.W.	FUEL GAS	BOILER FEED WATER MAKEUP	COOLING WATER MAKEUP	RAW WATER SUPPLY				
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016	3,830							
NITROGEN	N_2	28.016	462							
CARBON MONOXIDE	CO	28.01	520							
CARBON DIOXIDE	CO_2	44.01	532							
HYDROGEN SULFIDE	H_2S	34.076								
METHANOL	CH_3OH	32.042	99							
CARBONYL SULFIDE	COS	60.075								
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR		5,344							
WATER	H_2O	18.016								
TOTAL WET	lb MOLES/HR		5,443							
TOTAL	N POUNDS/HR			18	1,009	1,027				
MOLECULAR WEIGHT			11.3	18.016	18.016	18.016				
TEMPERATURE, °F			95	250	100	100				
PRESSURE, PSIA			20	1515	75	100				
MASCFH			2.07	-	-	-				
GPM			-	36	2,017	2,053				

TABLE 14
CONVERSION EFFICIENCY
KOPPERS-TOTZEK TO METHANOL

	10^6 BTU/HR INPUTS	10^6 BTU/HR OUTPUTS	EFFICIENCY, PERCENT
COAL-TO-MBG FACILITY	4575		
ELECTRIC POWER TO MBG FACILITY	1345.5		
SUBTOTAL	5920.5		
MBG PRODUCT		3196.2	
COAL TO MBG EFFICIENCY			69.9
OVERALL MBG EFFICIENCY			54.0
MBG TO ALTERNATE PRODUCTS	3196.2		
ELECTRIC POWER TO ALTERNATE PRODUCTS	468.9		
ALTERNATE PRODUCT		2510.7	
COAL-TO-ALTERNATE PRODUCT EFFICIENCY			54.9
OVERALL FACILITY EFFICIENCY			39.3

The gasoline synthesis unit is based on the Mobil M process which consists of two stages of fixed-bed catalytic reactors. In addition, the system also contains fractionation and alkylation units. The first stage of catalytic reactors has two parallel beds, while the second stage consist of six parallel beds.

Production capacity of the unit is 7350 barrels of gasoline per day, plus 1500 barrels of light hydrocarbon by-products (LPG) per day. To produce these products, 3143 short tons of methanol feed per day is required.

The biological treatment unit is designed to treat waste water from the gasoline synthesis process. Treatment capacity is 322 gallons per minute containing 3900 parts per million by weight (ppmw) of organic acids and phenols, and 5000 ppmw of acetone and gasoline.

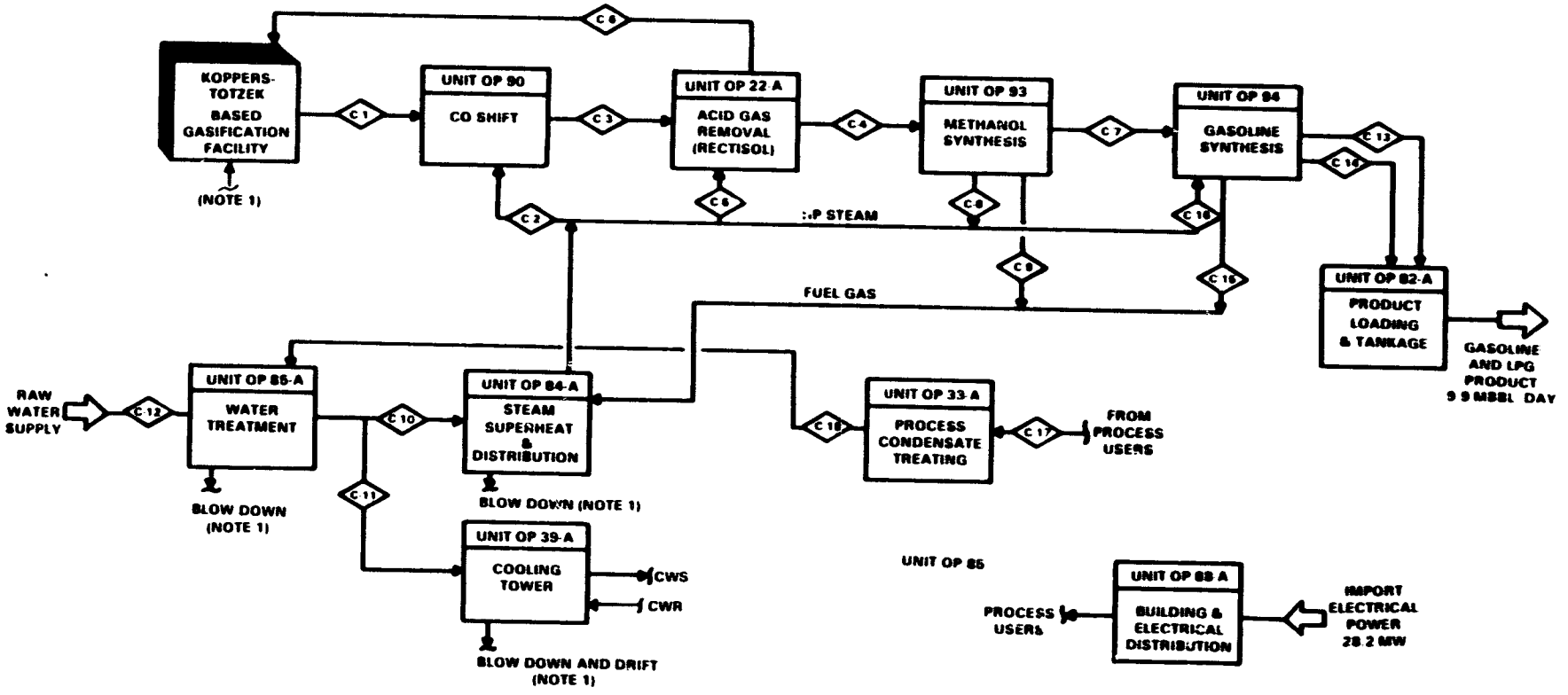
The material balance for gasoline production from MBG is presented in Table 15. The overall conversion efficiency for gasoline production is 30.9%. Detailed data are presented in Table 16.

Hydrogen

The block flow diagram for hydrogen production from MBG is presented in Sketch 290-514-SK-001D. The process units required for this add-on facility include acid gas removal, CO shift and hydrogen purification.

The Rectisol process is used for acid gas removal. Its requirements are the same as those for methanol production.

The CO shift unit utilizes a high temperature iron-chromium catalyst to shift stream composition from carbon monoxide to hydrogen. The inlet process stream is mixed with steam to produce a steam-to-dry gas ratio of 2.0. This stream is preheated to 600^{o}F and passed through two fixed-bed reactors in series. The inlet gas stream hydrogen-to-carbon monoxide ratio is 0.64. 14,460 moles of CO are shifted to produce a hydrogen-to-carbon monoxide ratio of 40.14 in the outlet stream.



NOTES: (1) BLOWDOWN IS ROUTED TO BASE PLANT FOR TREATMENT

BDM / MITTELHAUSER corporation	
NASA CONTRACT NAS8-33824	COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
514	ALTERNATE PRODUCT ANALYSIS KOPPERS-TOTZEK ADD ON CASE GASOLINE PRODUCTION MODULE BLOCK FLOW DIAGRAM
MDD/JVF	
290-514-SK-001C	0

TABLE 15
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE 1-C
KOPPERS-TOTZEK BASE TO GASOLINE

STREAM NUMBER			C-1	C-2	C-3	C-4	C-5	C-6	C-7	C-8
COMPONENT (COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)	SYMBOL	M.W.	RAW GAS FEED	STEAM TO CO SHIFT	AGR FEED	METHANOL SYNTHESIS FEED	STEAM TO AGR	ACID GAS	GASOLINE SYNTHESIS FEED (METHANOL)	STEAM FROM METHANOL SYNTHESIS
HYDROGEN	H_2	2.016	10,985		20,090	20,082		8	-	
NITROGEN	N_2	28.016	462		462	462		-	-	
CARBON MONOXIDE	CO	28.01	17,514		7,909	7,894		15	-	
CARBON DIOXIDE	CO_2	44.01	1,617		10,722	1,050		9,672	17	
HYDROGEN SULFIDE	H_2S	34.076	0.13		-	-		0.13	-	
METHANOL	CH_3OH	32.042	-		-	-		-	7,776	
CARBONYL SULFIDE	COS	60.075	0.60		0.60			0.60		
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR		30,079		39,184	29,688		9,696	7,793	
WATER	H_2O	18.016	6.1	15,087	65			65	501	
TOTAL WET	lb MOLES/HR		30,085	15,087	39,249	29,488		9,761	8,294	
TOTAL	lb POUNDS/HR		582.966	271.807	748.063	320.750	271.8	461.141	258.933	341.6
MOLECULAR WEIGHT			19.4	18.0	19.9	10.9	18.016	43.7	31.2	18.016
TEMPERATURE, °F			60	900	100	75	900	100	95	596
PRESSURE, PSIA			615	1500	575	500	1500	575	20	1515
MMSCFD			11.417	-	14.895	11.191	-	3.704	3.148	-
GPM							-			-

TABLE 15 (Continued)
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE 1-C
KOPPERS-TOTZEK BASE TO GASOLINE

STREAM NUMBER			C-9	C-10	C-11	C-12	C-13	C-14	C-15	C 16
STREAM NAME / COMPONENT	SYMBOL	M.W.	FUEL GAS	BOILER FEED WATER MAKEUP	COOLING WATER MAKEUP	RAW WATER SUPPLY	GASOLINE PRODUCT	LPG PRODUCT	FUEL GAS FROM GASOLINE SYNTHESIS	STEAM TO GASOLINE SYNTHESIS
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016	3,830							
NITROGEN	N_2	28.016	462							
CARBON MONOXIDE	CO	28.01	520							
CARBON DIOXIDE	CO_2	44.01	532							
HYDROGEN SULFIDE	H_2S	34.076								
METHANOL	CH_3OH	32.042	99							
CARBONYL SULFIDE	COS	60.075								
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052						110		
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12						174		
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146					44			
PENTENE	C_5H_{10}	70.13								
GASOLINE	C_{6+}	118*					666			
TOTAL DRY	lb MOLES/HR		5,346				710	284	57	
WATER	H_2O	18.016					-			
TOTAL GAS	lb MOLES/HR		5,443				710	284	57	
TOTAL	M POUNDS/HR			168.0	1,009	1,027	81.76	15.0	1.08	161.1
MOLECULAR WEIGHT			11.3	18.016	18.016	18.016	115.2	52.7	19.0	18.016
TEMPERATURE, °F			95	250	100	100	110	110	110	
PRESSURE, PSIA			20	1515	75	100	150	150	150	
MSCFH			2.07	-	-	-	-	-	0.02	-
GPH			-	336	2,017	2,053	235	54	-	322

*Average

TABLE 15 (Continued)
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE 1-C
KOPPERS-TOTZEK BASE TO GASOLINE

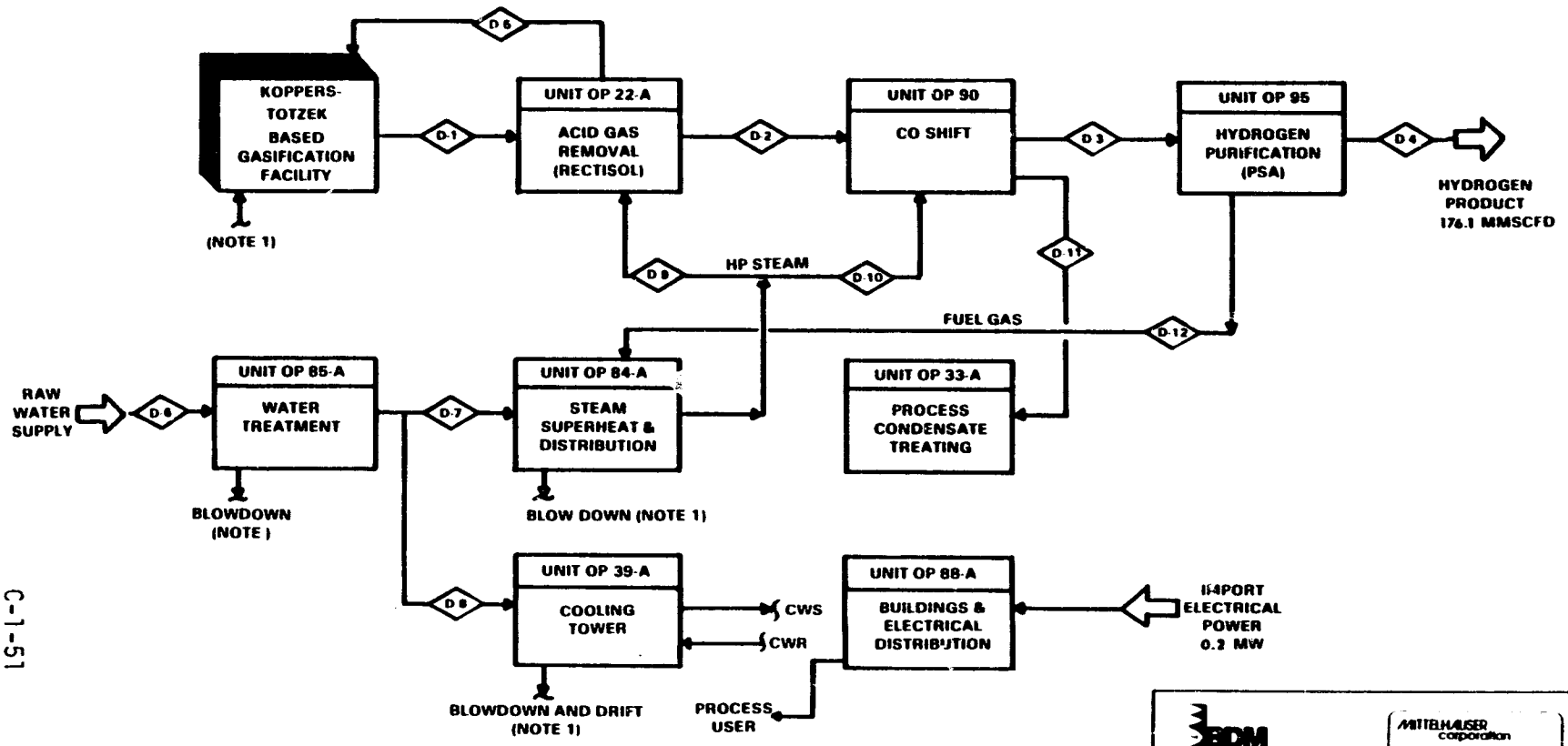
STREAM NUMBER			C-17	C-18	◇	◇	◇	◇	◇	◇
COMPONENT	SYMBOL	M.W.	PROCESS CONDENSATE	TREATED PROCESS CONDENSATE						
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016								
NITROGEN	N_2	28.016								
CARBON MONOXIDE	CO	28.01								
CARBON DIOXIDE	CO_2	44.01								
HYDROGEN SULFIDE	H_2S	34.076								
METHANOL	CH_3OH	32.042								
CARBONYL SULFIDE	COS	60.075								
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR									
WATER	H_2O	18.016	9,444	8,326						
TOTAL WET	lb MOLES/HR		9,444	8,326						
TOTAL	M POUNDS/HR		170.14	150.0						
MOLECULAR WEIGHT			18.016	18.016						
TEMPERATURE, °F			100	100						
PRESSURE, PSIA			65	65						
MSCFH			-	-						
GPM			340	300						

TABLE 16
CONVERSION EFFICIENCY
KOPPERS-TOTZEK TO GASOLINE

	10^6 BTU/HR INPUTS	10^6 BTU/HR OUTPUTS	EFFICIENCY, PERCENT
COAL-TO-MBG FACILITY	4575		
ELECTRIC POWER TO MBG FACILITY	1345.5		
SUBTOTAL	5920.5		
MBG PRODUCT		3196.2	
COAL TO MBG EFFICIENCY			69.9
OVERALL MBG EFFICIENCY			54.0
MBG TO ALTERNATE PRODUCTS	3196.2		
ELECTRIC POWER TO ALTERNATE PRODUCTS	385.4		
ALTERNATE PRODUCT		1947.0	
COAL-TO-ALTERNATE PRODUCT EFFICIENCY			42.6
OVERALL FACILITY EFFICIENCY			30.9

The hydrogen purification process is based on a pressure swing absorption process licensed by Union Carbide. This unit is designed to recover 99.99% pure hydrogen from an inlet stream of 9950 moles per hour of 60.7% hydrogen by volume. The purge gas hydrogen loss is 20% of the inlet stream hydrogen.

The material balance for hydrogen production from MBG is given in Table 17. The overall conversion efficiency for hydrogen production is 40.0%. Detailed data are presented in Table 18.



NOTES: (1) BLOWDOWN IS ROUTED TO BASE PLANT FOR TREATMENT.

BDM / MITTELHAUSER corporation	
NASA CONTRACT NAS8-33824	COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.1.4	ALTERNATE PRODUCT ANALYSIS KOPPERS-TOTZEK ADD-ON CASE HYDROGEN PRODUCTION MODULE BLOCK FLOW DIAGRAM
MDD/	
MDD/JVF	
290-514-SK-001D	0 — 7/15/80

TABLE 17
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE 1-D
KOPPERS-TOTZEK BASE TO HYDROGEN

STREAM NUMBER			D-1	D-2	D-3	D-4	D-5	D-6	D-7	D-8
COMPONENT (STREAM NAME)	SYMBOL	M.W.	RAW GAS FEED	CO SHIFT FEED	PSA FEED	HYDROGEN PRODUCT	ACID GAS	RAW WATER SUPPLY	BOILER FEED WATER MAKEUP	COOLING WATER MAKEUP
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016	9,744	9,705	24,166	19,333	39			
NITROGEN	N_2	28.016	410	410	410	-	-			
CARBON MONOXIDE	CO	28.01	15,092	15,063	602	2	29			
CARBON DIOXIDE	CO_2	44.01	1,434	156	14,617		1,278			
HYDROGEN SULFIDE	H_2S	34.076	0.12				0.12			
METHANOL	CH_3OH	32.042								
CARBONYL SULFIDE	COS	60.075	0.53				0.53			
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6									
TOTAL DRY	lb MOLES/HR		26,681	25,334	39,795	19,335	1,347			
WATER	H_2O	18.016	5.4	-	76	-	5.4			
TOTAL WET	lb MOLES/HR		26,686	25,334	39,871	19,335	1,352			
TOTAL	M POUNDS/HR		517.101	459.832	271.731	39.031	57.172	762.5	307	455.5
MOLECULAR WEIGHT			19.4	18.2	18.1	2.0	42.3	18.0	18.0	18.0
TEMPERATURE, °F			60	75	100	100	75	100	250	75
PRESSURE, PSIA			615	540	500	300	540	100	1515	75
MISCFH			10.127	9.614	15.131	7.338	0.513	-	-	-
GPM			-	-	-	-	-	1525	614	911

TABLE 17 (Continued)
MATERIAL BALANCE
ALTERNATE PRODUCTS ANALYSIS
CASE I-D
KOPPERS-TOTZEK BASE TO HYDROGEN

STREAM NUMBER			D-10	D-11	D-12					
COMPONENT / STREAM NAME	SYMBOL	M.W.	STEAM TO CO SHIFT	COOLING WATER USERS	LBG FUEL					
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016			4,833					
NITROGEN	N_2	28.016			410					
CARBON MONOXIDE	CO	28.01			600					
CARBON DIOXIDE	CO_2	44.01			14,617					
HYDROGEN SULFIDE	H_2S	34.076								
METHANOL	CH_3OH	32.042								
CARBONYL SULFIDE	COS	60.075								
METHANE	CH_4	16.042								
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR				20,384					
WATER	H_2O	18.016	50,668		76					
TOTAL WET	lb MOLES/HR		50,668		20,460					
TOTAL	M POUNDS/HR		912.835		682.699					
MOLECULAR WEIGHT			18.0		33.4					
TEMPERATURE, °F			596		100					
PRESSURE, PSIA			1500		20					
PHASE			-		7.765					
GPM										

TABLE 18

CONVERSION EFFICIENCY
KOPPERS-TOTZEK TO HYDROGEN

	10^6 BTU/HR		EFFICIENCY,
	INPUTS	OUTPUTS	PERCENT
COAL-TO-IBG FACILITY	4575		
ELECTRIC POWER TO IBG FACILITY	1345.5		
SUBTOTAL	5920.5		
IBG PRODUCT		3196.2	
COAL TO IBG EFFICIENCY			69.9
OVERALL IBG EFFICIENCY			54.0
IBG TO ALTERNATE PRODUCTS	3196.2		
ELECTRIC POWER TO ALTERNATE PRODUCTS	24.6		
ALTERNATE PRODUCT		2377.4	
COAL-TO-ALTERNATE PRODUCT EFFICIENCY			52.0
OVERALL FACILITY EFFICIENCY			40.0

COST ANALYSIS AND RESULTS

Costs were determined for two separate areas: initial capital and operating costs. The methodologies used for both areas, as well as the results of the analysis are presented in this section.

Approach To Capital Costs

The first step in the analysis is to select a base or reference system from the design data base. There are two criteria that were adhered to in this selection process. First, the capacity of the reference system should be as close as possible to current capacity needs. This minimizes errors associated with the use of an incorrect cost-capacity exponent.

The second criteria is that the reference system configuration should closely match current design needs. Any design adjustment required by a mismatch should be straightforward.

The second step in the analysis is to select the appropriate size parameter with which the reference unit can be scaled. The parameters chosen for each of the required reference systems are presented in Table 19.

Once the appropriate reference systems were selected and the sizing parameters identified, the costs were scaled using a cost-capacity exponent of 0.6. Final cost adjustments were then made to account for escalation to January of 1980, indirect costs not included in the reference cost, and the engineering and design costs.

Two examples of the use of this methodology are presented below.

Example 1 - Acid Gas Removal Unit Cost Adjustment

System:	Rectisol
Reference:	FE-2542-10 "Conoco Demonstration Plant For High BTG Gas,"
Date:	April, 1978

Several subsystems in this design were needed for purification of Lurgi-derived gas that are not required for a K-T system. These include Naphtha Recovery (there is no naphtha in K-T MBG) and CO_2 off gas compression (there are no lock hoppers in the K-T gasifier). The cost of this equipment was deducted from overall equipment cost and the system cost estimate was then rebuilt using the same techniques as in the reference. This

TABLE 19

SCALING PARAMETERS SELECTION

SHIFT CONVERSION, USE MOLS/HR CO SHIFTED.
ACID GAS REMOVAL, USE MOLS/HR ACID GAS (H_2S + COS + CO_2) REMOVED.
METHANATION, USE MOLS/HR CH_4 PRODUCED.
METHANOL, USE MOLS/HR CH_3OH PRODUCED.
GASOLINE SYNTHESIS, USE MOLS/HR CH_3OH FED TO SYSTEM.
HYDROGEN PURIFICATION, USE MOLS/HR PRODUCT H_2 PRODUCED.

resulted in a reference system capital reduction of 15 percent.

The utilities required by these subsystems were deducted from the totals for the AGR system. This resulted in a reduction of system steam comsumption of 20 percent.

Cost of the system is obtained by scaling on moles per hour of H_2S, COS and CO_2 removed. Cost adjustments including scaling, escalation, indirect factor, and engineering and design costs result in

$$\text{Single Module Cost} = \$32.27 \times 10^6 \times (13{,}556/13{,}827)^{0.6} \times 1.22 \times 1.241 \times 1.08$$

$$= \$52.14 \times 10^6$$

where $\$32.27 \times 10^6$ is adjusted cost of reference system,

13,556 is moles per hour of acid gas removed by add-on facility system,

13,827 is moles per hour of acid gas removed by reference system,

0.6 is scale factor,

1.22 is escalation factor,

1.241 is indirect factor,

and 1.08 is engineering and design factor.

Example 2 - CO Shift Unit

System:	Shift Conversion
Reference:	FE-2240-31 "Factored Estimates for Eastern Coal Commercial Concepts"
Date:	January, 1976

This system was selected because it has a reasonable configuration that has a scale close to the current design and uses proven concepts and equipment. Cost of the system is obtained by scaling on moles per hour of CO shifted.

Cost adjustments result in

$$\text{Single Module Costs} = \$14.5 \times 10^6 \times \left(\frac{9,291}{7,141}\right)^{0.6} \times 1.30 \times 1.00 \times 1.08$$

$$= \$23.84 \times 10^6$$

where $\$14.5 \times 10^6$ is cost of reference system,

9,291 is moles per hour of CO shifted by add-on facility system,

7,141 is moles per hour of CO shifted by reference system,

0.6 is scale factor,

1.30 is escalation factor,

1.00 is indirect factor (none),

1.08 is engineering and design factor.

Capital Cost Results

The capital costs for the four alternate-product modules for the K-T add-on facilities are detailed in Tables 20, 21, 22 and 23. Costs are given in 1980 dollars for each system, both onsite and offsite, as well as project contingency, owner's engineering and G&A, and the contractors fee.

Approach To Operating Costs

Operating costs were calculated based on the K-T MBG facility where possible, and on reference literature otherwise. Operating labor is linearly scaled from system labor costs found in the design data base. Power consumption is calculated by means of a utility balance. The maintenance labor and supplies are calculated as a percentage of system capital investment. These percentages amount to 3% for onsites and 1% for offsites. Supervisory costs are added according to engineering judgment. Other labor costs are calculated as percentages of operating and maintenance labor, or as a percentage of subtotal operating costs.

Operating Cost Results

The first year operating costs for the four alternate-product modules for the K-T add-on facilities are detailed in Tables 24, 25, 26 and 27. Costs are given in 1980 dollars on a facility basis. Costs are broken down in

terms of operating labor, operating supplies, maintenance labor, maintenance supplies, supervision, general plant staff, catalyst and chemical makeup, electric power, and administration and general overhead.

TABLE 20

CAPITAL COSTS FOR METHANE KOPPERS-TOTZEK, ADD-ON FACILITY

	SYSTEM DESCRIPTION	MILLIONS OF 1980 DOLLARS PER MODULE	PER FACILITY
90	SHIFT CONVERSION	23.84	95.36
22-A	ACID GAS REMOVAL (RECTISOL)	44.70	178.80
91	METHANATION	26.89	107.56
92	DRYING	.36	1.42
	SUBTOTAL ONSITES	95.79	383.14
39-A	COOLING TOWER	3.96	15.84
84-A	STEAM SUPERHEATERS	3.62	14.46
85-A	RAW WATER TREATMENT	4.70	18.78
87/88	GENERAL AND SUPPORT FACILITIES (INCLUDES TURBOGENERATOR SET)	7.50	30.02
	SUBTOTAL OFFSITES	19.78	79.10
	TOTAL SYSTEM CAPITAL INVESTMENT		462.24
	PROJECT CONTINGENCY		69.34
	OWNERS ENGINEERING, G&A		10.63
	CONTRACTOR'S FEE		21.26
	TOTAL FACILITY INVESTMENT		563.47

TABLE 21

CAPITAL COSTS FOR METHANOL
KOPPERS-TOTZEK, ADD-ON FACILITY

SYSTEM	DESCRIPTION	MILLIONS OF 1980 DOLLARS PER MODULE	MILLIONS OF 1980 DOLLARS PER FACILITY
90	SHIFT CONVERSION	23.55	94.21
22-A	ACID GAS REMOVAL (RECTISOL)	42.59	170.34
93	METHANOL SYNTHESIS	32.42	129.64
	SUBTOTAL ONSITES	98.56	394.19
39-A	COOLING TOWER	3.56	14.25
85-A	RAW WATER TREATMENT	4.23	16.92
87/88	GENERAL AND SUPPORT FACILITIES (INCLUDING METHANOL STORAGE-30 DAYS)	6.33	25.31
	SUBTOTAL OFFSITES	14.12	56.48
	TOTAL SYSTEM CAPITAL INVESTMENT		450.67
	PROJECT CONTINGENCY		67.60
	OWNERS ENGINEERING, G&A		10.37
	CONTRACTOR'S FEE		20.74
	TOTAL FACILITY INVESTMENT		549.37

TABLE 22

CAPITAL COSTS FOR GASOLINE KOPPERS-TOTZEK, ADD-ON FACILITY

SYSTEM DESCRIPTION	MILLIONS OF 1980 DOLLARS PER MODULE	MILLIONS OF 1980 DOLLARS PER FACILITY
ADD-ON FACILITIES FOR METHANOL	112.67	450.67
94 GASOLINE SYNTHESIS	53.35	213.40
37-A WASTE WATER BIOLOGICAL TREATMENT	5.62	22.50
ADDITIONAL GENERAL AND SUPPORT FACILITIES, INCLUDING GASOLINE AND BYPRODUCT STORAGE IN PLACE OF METHANOL STORAGE		2.60
TOTAL SYSTEM CAPITAL INVESTMENT		689.17
PROJECT CONTINGENCY		103.38
OWNERS ENGINEERING, G&A		15.85
CONTRACTOR'S FEE		31.70
TOTAL FACILITY INVESTMENT		840.10

TABLE 23

CAPITAL COSTS FOR HYDROGEN KOPPERS-TOTZEK, ADD-ON FACILITY

SYSTEM	DESCRIPTION	MILLIONS OF 1980 DOLLARS PER MODULE	PER FACILITY
22-A	ACID GAS REMOVAL (RECTISOL)	12.65	50.59
90	SHIFT CONVERSION	31.09	124.35
95	PRESSURE - SWING ADSORPTION	86.96	347.85
	SUBTOTAL ONSITES	130.70	522.79
39-A	COOLING TOWER	2.21	8.85
33-A	PROCESS CONDENSATE TREATMENT	.48	1.92
84-A	STEAM GENERATION	58.80	235.21
85-A	RAW WATER TREATMENT	3.58	14.32
87/88	GENERAL AND SUPPORT FACILITIES	6.07	24.28
	SUBTOTAL OFFSITES	71.14	284.58
	TOTAL SYSTEM CAPITAL INVESTMENT		807.37
	PROJECT CONTINGENCY		121.10
	OWNERS ENGINEERING, G&A		18.57
	CONTRACTOR'S FEE		37.14
	TOTAL FACILITY INVESTMENT		984.18

TABLE 24

OPERATING COSTS FOR METHANE KOPPERS-TOTZEK, ADD-ON FACILITY

CATEGORY	QUANTITY	MILLION OF 1980 DOLLARS PER FACILITY
OPERATING LABOR	26 PERSONS PER SHIFT PER FACILITY	2.74
OPERATING SUPPLIES	15% OF OPERATING LABOR	.41
MAINTENANCE		
TOTAL	3% OF ONSITE SYSTEM CAPITAL + 1% OF OFFSITE SYSTEM CAPITAL	
LABOR	40% OF TOTAL MAINTENANCE	4.91
SUPPLIES	60% OF TOTAL MAINTENANCE	7.37
SUPERVISION	3 ADDITIONAL SUPERVISORS	.14
GENERAL PLANT STAFF	30% OF OPERATING & MAINTENANCE LABOR	2.30
CATALYST & CHEMICAL MAKE-UP	SYSTEM REQUIREMENTS	8.80
ELECTRIC POWER	7227 KW CREDIT PER MODULE	(5.65)
ADMINISTRATION & GENERAL OVERHEAD	5% OF OPERATING COSTS	1.05
TOTAL ANNUAL ADD-ON OPERATING EXPENSES		22.07

TABLE 25

OPERATING COSTS FOR METHANOL KOPPERS-TOTZEK, ADD-ON FACILITY

CATEGORY	QUANTITY	MILLIONS OF 1980 DOLLARS PER FACILITY
OPERATING LABOR	25 PERSONS PER SHIFT PER FACILITY	2.67
OPERATING SUPPLIES	15% OF OPERATING LABOR	.40
MAINTENANCE		
TOTAL	30% OF ONSITE SYSTEM CAPITAL + 1% OF OFFSITE SYSTEM CAPITAL	
LABOR	40% OF TOTAL MAINTENANCE	4.96
SUPPLIES	60% OF TOTAL MAINTENANCE	7.44
SUPERVISION	3 ADDITIONAL SUPERVISORS	.14
GENERAL PLANT STAFF	30% OF OPERATING & MAINTENANCE LABOR	2.29
CATALYST & CHEMICAL MAKEUP	SYSTEM REQUIREMENTS	3.15
ELECTRIC POWER	34,344 KW PER MODULE	26.83
ADMINISTRATION & GENERAL OVERHEAD	5% OF OPERATING COSTS	2.39
TOTAL ANNUAL ADD-ON OPERATING EXPENSES		50.27

TABLE 26

OPERATING COSTS FOR GASOLINE KOPPERS-TOTZEK, ADD-ON FACILITY

CATEGORY	QUANTITY	MILLIONS OF 1980 DOLLARS PER FACILITY
OPERATING LABOR	39 PERSONS PER SHIFT PER FACILITY	4.08
OPERATING SUPPLIES	15% OF OPERATING LABOR	.61
MAINTENANCE		
TOTAL	3% OF ONSITE SYSTEM CAPITAL + 1% OF OFFSITE SYSTEM CAPITAL	
LABOR	40% OF TOTAL MAINTENANCE	8.05
SUPPLIES	60% OF TOTAL MAINTENANCE	12.07
SUPERVISION	4 ADDITIONAL SUPERVISORS	.18
GENERAL PLANT STAFF	30% OF OPERATING & MAINTENANCE LABOR	3.64
CATALYST & CHEMICAL MAKEUP	SYSTEM REQUIREMENTS	3.51
ELECTRIC POWER	28,233 KW PER MODULE	.10
ADMINISTRATION & GENERAL OVERHEAD	5% OF OPERATING COSTS	2.71
TOTAL ANNUAL ADD-ON OPERATING EXPENSES		56.95

TABLE 27

OPERATING COSTS FOR HYDROGEN KOPPERS-TOTZEK, ADD-ON FACILITY

CATEGORY	QUANTITY	MILLIONS OF 1980 DOLLARS PER FACILITY
OPERATING LABOR	45 PERSONS PER SHIFT PER FACILITY	4.78
OPERATING SUPPLIES	15% OF OPERATING LABOR	.72
MAINTENANCE		
TOTAL	3% OF ONSITE SYSTEM CAPITAL + 1% OF OFFSITE SYSTEM CAPITAL	
LABOR	40% OF TOTAL MAINTENANCE	7.41
SUPPLIES	60% OF TOTAL MAINTENANCE	11.12
SUPERVISION	3 ADDITIONAL SUPERVISORS	.14
GENERAL PLANT STAFF	30% OF OPERATING & MAINTENANCE LABOR	3.66
CATALYST & CHEMICAL MAKEUP	SYSTEM REQUIREMENTS	2.31
ELECTRIC POWER	1800 KW PER MODULE	1.41
ADMINISTRATION & GENERAL OVERHEAD	5% OF OPERATING COSTS	1.30
TOTAL ANNUAL ADD-ON OPERATING EXPENSES		27.35

ALTERNATE GASIFIERS

The original scope of this task called for a qualitative assessment of the remaining gasifier technologies (Lurgi, BGC Lurgi, and Babcock and Wilcox) as to the alternate product each was best suited. This scope has been modified, in that the Lurgi technology has been used to design and cost an add-on facility to produce methane and methanol, and will be reported in a later report. The following is a description of the B&W Lurgi gasifier technologies in regard to alternate products.

Babcock and Wilcox

The B&W gasifier is an entrained flow gasifier, similar to the Texaco and K-T gasifiers. The typical gas product is a medium BTU gas (300 BTU/SCF) consisting primarily of H_2, CO, CO_2, and sulfur compounds. There is essentially no methane, or ammonia or tars formed in the gasification process. The H_2/CO ratio of the gas is approximately 0.43. The B&W product gas is best suited for MBG or as synthesis gas for methanol, ammonia, hydrogen, etc., rather than synthetic natural gas (SNG) production as compared to the Lurgi or BGC/Lurgi cases.

BGC Lurgi

The BGC Lurgi slaggging gasifier is a moving bed gasifier similar in operation and product gas yield to the Lurgi (dry bottom) gasifier. The product gas contains significant amounts of methane, tar, and oils and has a higher heating value (HHV) of greater than 300 BTU/SCF. The clean gas product has an H_2/CO ratio in the range of 2.0 to 2.3. The methane content of the gas (9.0%) and the high H_2/CO ratio make SNG production a significant candidate for this gasifier technology. A difference between the BGC Lurgi and the Lurgi (dry bottom) gasifiers is that the Lurgi gasifier requires more steam injection in the gasification process. This results in a higher H_2/CO ratio (2.5) and consequently a stream more suitable for SNG production.

BIBLIOGRAPHY

FE-2240-101, TECHNICAL REPORT FOR JOINT DOE-GRI COAL GASIFICATION PROGRAM (FINAL REPORT), NOVEMBER, 1978

FE-2240-31 FACTORED ESTIMATES FOR EASTERN COAL COMMERCIAL CONCEPTS (INTERIM REPORT), C. F. BRAUN AND COMPANY, SEPTEMBER 1978

FE-2240-49 COMPARISON OF ACID GAS REMOVAL SYSTEMS (FINAL REPORT), C. F. BRAUN AND COMPANY, APRIL, 1978

FE-2447-13 RESEARCH GUIDANCE STUDIES TO ASSESS GASOLINE FROM COAL BY METHANOL-TO-GASOLINE AND SASOL-TYPE FISCHER-TROPSCH TECHNOLOGIES (FINAL REPORT), MOBIL RESEARCH AND DEVELOPMENT CORPORATION, AUGUST, 1978

APPENDIX C-2
METHANE/METHANOL PRODUCTION:
LURGI GASIFICATION BASED PLANT

TABLE OF CONTENTS

LIST OF TABLES

LIST OF DRAWINGS

1.0 INTRODUCTION

This report presents the results and a discussion of the process design of units necessary to produce methanol and/or methane from Medium BTU Gas (MBG) produced in a Lurgi-based gasification facility. The purpose of this study was twofold: to provide system level engineering design and evaluation of a Lurgi-based gasification facility producing methane or methane-methanol as an alternate to MBG; and, secondly, to determine the capital and operating costs associated with each alternate product, which would be used to develop relative alternate product costs.

The alternate product systems described in this report have been designed as add-on units to the Lurgi facility definition design that has been reported to the Marshall Space Flight Center (MSFC) in an earlier report[1] on September 3, 1980. The same level of design and methodology that was used in the Koppers-Totzek and Texaco-based alternate product analysis have been used in this study.

Each of the case designs for the production of methane or methanol from Lurgi MBG is intended as an extension of previous contract subtasks. For each case of the system design, available technologies were identified and comparative trade-off studies preformed to select the preferred technologies for design. Each system design was based on published data, engineering experiences and judgment, and design calculations specific to this study.

There are two cases of alternate product analysis presented in this report. Lurgi-derived MBG contains an appreciable amount of methane (8.97 per cent) due to the Lurgi gasification process. Case II-A is the study designated to produce methane from MBG. Case II-B is the study designated for the co-production of methane/methanol. The co-production case was prompted by two factors. First, methanol production results in an

off-gas suitable for a methanation reaction. Second, since TVA has an apparent market for methane, it was decided that it was more economically feasible to synthesize additional methane rather than breakdown the existing methane into a synthesis gas.

System level costs, based on cost-versus-capacity factoring, are presented on a modular basis. Facility costs were determined based upon the study results and module costs.

The discussion that follows describes the process design and results of this analysis at the module level.

2.0 SUMMARY

A definition level process design to produce methane or methanol has been prepared based on Lurgi-derived MBG. The process design has been performed on module level capacities. The Lurgi based module receives 5,000 tons/day sized coal at the gasifiers and produces 290 MM SCF/Day of MBG with a heating value of 380 BTU/SCF. This net quantity of MBG is converted into methane or methanol. Detailed material balances are included in a later section of this report, along with their associated overall block flow diagrams. However, the key material input and outputs are summarized in Table 2-1.

The facility costs to produce methane or methane/methanol, presented in Tables 2-2 through 2-5 respectively, present the costs of a facility (4 modules) necessary to produce 100% of the alternate product.

Based on the data presented in these tables and the methodology described in BDM/W-80-258-TR-RV2, "Cost Estimation and Economic Evaluation Methodology," the 1980 cost of product is $7.69/MM Btu for methane and $6.81/MM Btu for a co-product slate of methanol and methane.

The following sections of this report provide more details concerning the process design of key elements of the alternate product facility.

TABLE 2-1. KEY MATERIAL INPUT AND PRODUCTS

	LURGI-DERIVED MBG FEED GAS	PRODUCTS	
		CASE II-A METHANE	CASE II-B METHANE/METHANOL
PRODUCT COMPOSITION, MOLE %			
H_2	46.77	4.96	9.12 / -
N_2	0.44	1.59	2.73 / -
CO	17.09	0.05	NIL / -
CO_2	26.02	0.89	0.05 / 0.16
SULFUR	200 PPMV	NIL	NIL / NIL
CH_4	8.97	92.50	88.08 / 0.04
C_2^+	0.68	-	- / -
CH_3OH	-	-	- 97.64
H_2O	114 PPMV	100 PPMV	160 PPMV / 2.16
MMSCFD - FEED GAS	289.5	-	- / -
MMSCFD - METHANE	(26.0)*	75.6	46.8 / (0.08)*
M BARREL/DAY - METHANOL	-	-	- / 14.25
GAS HHV - BTU/SCF	308	950.3	918.9 / -
IMPORT POWER - KW	29.4	20.2	24.1
IMPORT WATER - M GAL/DAY	2,999.5	411.8	1,137.6

NOTE: Results are representative of one module.

*Included in feed gas.

TABLE 2-2. TOTAL CAPITAL REQUIREMENTS -- ADD-ON METHANATION FACILITY

UNIT OPERATION	CAPITAL INVESTMENT MILLION DOLLARS
SHIFT CONVERSION	20.42
ACID GAS REMOVAL	160.26
METHANATION	98.44
GAS DRYING	1.60
COOLING WATER	53.12
STEAM SUPERHEAT	18.79
RAW WATER TREATING	14.55
SOLIDS TREATING	23.80
BUILDINGS AND ELECTRICAL DISTRIBUTION	12.12
SYSTEM CAPITAL INVESTMENT (ADD-ON METHANATION FACILITY)	403.10
PROJECT CONTINGENCY	60.47
OWNERS ENGINEERING & G AND A	9.64
CONTRACTOR'S FEE	18.54
	491.75
SYSTEM CAPITAL INVESTMENT (LURGI BASE FACILITY)	1,879.00
TOTAL FACILITY INVESTMENT	2,370.75
ROYALTIES	11.85
AFUDC	506.31
START-UP AND TESTING	483.09*
TOTAL DEPRECIABLE INVESTMENT	3,372.00
NON-DEPRECIABLE INVESTMENT	101.16
TOTAL CAPITAL REQUIREMENTS	3,473.16

*Includes coal used during start-up and testing.

TABLE 2-3. OPERATING AND MAINTENANCE COSTS -- ADD-ON METHANATION FACILITY

	MILLION DOLLARS
OPERATING LABOR	1.88
OPERATING SUPPLIES	0.28
MAINTENANCE LABOR	1.97
MAINTENANCE SUPPLIES	2.95
SUPERVISION	0.14
GENERAL PLANT STAFF	1.20
CATALYST AND CHEMICALS	6.76
ELECTRIC POWER	12.30
ADMINISTRATION AND GENERAL OVERHEAD	1.04
ANNUAL OPERATING COSTS (ADD-ON METHANE FACILITY)	28.52
ANNUAL OPERATING COST (LURGI BASE PLANT)	366.38
TOTAL ANNUAL OPERATING COSTS	429.99

TABLE 2-4. TOTAL CAPITAL REQUIREMENTS -- ADD-ON METHANE/METHANOL FACILITY

UNIT OPERATION	CAPITAL INVESTMENT MILLION DOLLARS
METHANOL SYNTHESIS	100.76
ACID GAS REMOVAL	153.12
METHANATION	49.16
COMPRESSION	23.36
GAS DRYING	1.16
CO_2 REMOVAL	40.20
COOLING WATER	70.11
STEAM SUPERHEAT	14.48
RAW WATER TREATING	27.96
BY-PRODUCT STORAGE	15.24
BUILDINGS AND ELECTRICAL DISTRIBUTION	15.36
SYSTEM CAPITAL INVESTMENT (ADD-ON METHANATION/METHANOL FACILITY)	510.91
PROJECT CONTINGENCY	76.64
OWNERS ENGINEERING & G AND A	12.22
CONTRACTOR'S FEE	23.50
	623.27
SYSTEM CAPITAL INVESTMENT (LURGI BASE FACILITY)	1,879.00
TOTAL FACILITY INVESTMENT	2,502.27
ROYALTIES	12.51
AFUDC	534.38
START-UP AND TESTING	408.27*
TOTAL DEPRECIABLE INVESTMENT	3,457.43
NON-DEPRECIABLE INVESTMENT	102.16
	3,559.59

*Includes coal used during start-up and testing.

TABLE 2-5. OPERATING AND MAINTENANCE COSTS -- ADD-ON METHANE/METHANOL FACILITY

	MILLION DOLLARS
OPERATING	2.43
OPERATING SUPPLIES	0.36
MAINTENANCE LABOR	2.49
MAINTENANCE SUPPLIES	3.74
SUPERVISION	0.14
GENERAL PLANT STAFF	1.52
CATALYST AND CHEMICALS	5.32
ELECTRIC POWER	15.13
ADMINISTRATION AND GENERAL OVERHEAD	1.29
ANNUAL OPERATING COSTS (ADD-ON METHANE FACILITY)	32.42
ANNUAL OPERATING COST (LURGI BASE PLANT)	366.38
TOTAL ANNUAL OPERATING COSTS	398.80

3.0 DESIGN BASIS AND SCOPE

3.1 Background

The alternate product analyses have been predicated on a desire to provide the TVA Gasification Project Team sufficient data on cost and process requirements to make a decision on whether to add alternate product capability to their proposed gasification facility. Earlier team work[2] on alternate product analysis considered a wide spectrum of alternates (methane, methanol, hydrogen, and gasoline) derived from MBG produced in a Koppers-Totzek or Texaco-based gasification facility.

The objective of this study was to provide system level definition of a facility that produced methanol and/or methane from MBG derived from a Lurgi-based gasification facility.

3.2 Design Basis

The alternate project analysis presented in this report has been performed at the same level of design as the previous work on Koppers-Totzek and Texaco to provide a similar base for comparison. The Lurgi facility definition design has been designed in accordance with the TVA Design Criteria, April 1980. This design and the results have been submitted to MSFC in an earlier report.

As in the earlier alternate product analysis, BDM/H-80-481-TR, the alternate product units have been considered "add-on" units with minimal integration to the base facility. The base facility is considered to be the Lurgi-based MBG facility.

Several engineering design studies have been performed and reported in open literature. These sources provide an excellent basis for the

factored systems designs reported herein for the CO shift unit[3], the Rectisol unit[4], the methanol synthesis unit[5], and the methanation unit[3]. These designs were appropriately scaled for our feed quantities, as was done in the earlier alternate product analysis, and used to develop the expected product yields and utility requirements, as well as systems costs. The module systems costs were then used to develop the facility level capital and operating costs.

The net MBG from one Lurgi module was used as feed to the alternate product facilities. Fuel gas required for internal consumption (steam super heat) in the alternate facilities was taken from the net methane product. This was done because (1) no apparent cost savings were observed for reduced through-put in the alternate product facilities and (2) the HHV of the methane product requires less gas volume and utilizes a high efficiency fired heater.

3.3 Approach to Design

The approach taken in this study to develop a process design of the facilities necessary to produce methane/methanol from Lurgi-derived MBG was essentially the same as that in the K-T and Texaco alternate product analysis done earlier[2]. In any continuing process engineering study, earlier tradeoff studies and design results are utilized in the later studies which tend to produce a more optimum design.

The facilities to produce methanol and/or methane from Lurgi-derived MBG have been designed as an add-on facility to the Lurgi base facility*. There was no intent to integrate the two facilities, which resulted in duplication of some units, both in the process units and in the off-site units. However, observations resulting from the K-T and Texaco alternate product analysis indicated that a more realistically economic

*The Lurgi base facility is considered to be the stand-alone Lurgi-based gasification facility that produces MBG as its product gas.

design could be achieved by processing small waste water and solid slurry streams in the Lurgi base facility. These considerations are identified and discussed in later sections of this report.

The purpose of this alternate product analysis was twofold: first, to provide system level engineering design and evaluation of a Lurgi-based gasification facility producing methane and/or methanol as an alternate to MBG; and secondly, to determine the capital and operating costs associated with each alternate product.

An add-on, modular approach was used as the design basis for design and costing of each alternate product. Each alternate product facility was based on the net output of MBG, derived from 5,000 tons per day coal feed, as defined in the Lurgi facility definition design[1]. The facility level costs and capacities were then determined based on a four module facility.

The trade-off studies performed in the K-T and Texaco alternate product analysis concerning each of the unit operations necessary to produce methanol and/or methane are still valid for this study.

The processing scheme selected includes CO shift to adjust the H_2/CO ratio to the value necessary for the catalytic reaction, a selective Rectisol unit to remove CO_2 and sulfur compounds and the catalytic reaction to produce methane or methanol. Also, a gas drying unit is provided for the methane product to meet water content requirements.

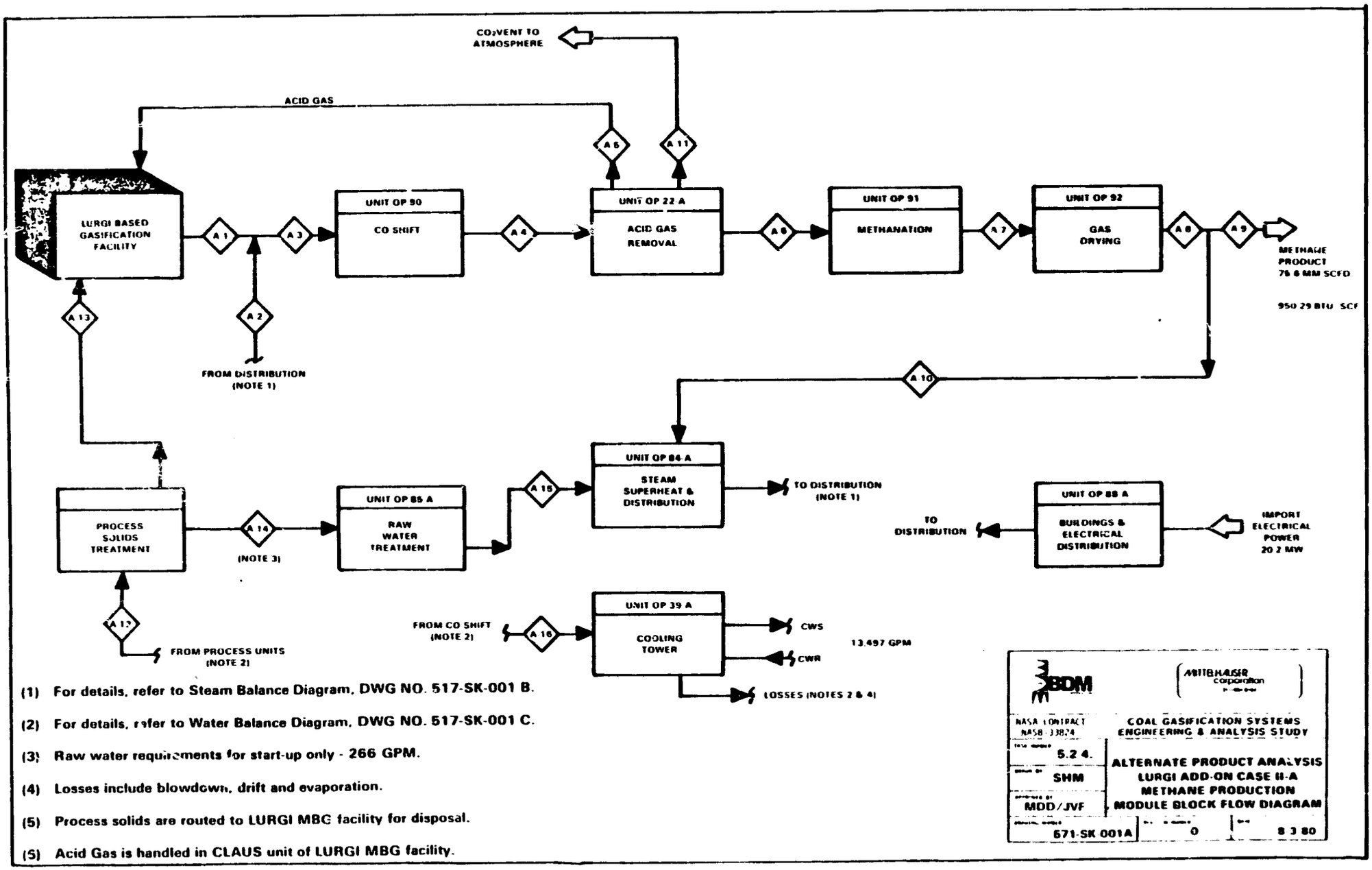
Furthermore, the steam and water requirements of the alternate product facility are met within the alternate facility boundary.

3.3.1 Lurgi Base to Methane - Case II-A

Case II-A was designated as the analysis of facilities to produce methane from Lurgi-derived MBG. The process units necessary to produce methane are shown on drawing 571-SK-001A, Methane Production Module Block Flow Diagram. The associated material balance is provided in Table 3-1. The process sequence is similar to that provided in the Koppers-Totzek alternate product analysis.

Lurgi-derived MBG at 100°F and 615 PSIA is mixed with HP steam and fed to a CO shift conversion unit. CO shift effluent is fed to a selective rectisol unit, which is used to remove CO_2 to 1.0 per cent of the treated gas and essentially all of the sulfur compounds. The shifted, surfur free synthesis gas is then fed to the methanation unit where 99.9 per cent of the CO, 68.8 per cent of the CO_2, and 100 per cent of the ethane/ethylene components are converted to methane. The final step in the process is a TEG drying unit to dry the product methane to pipeline specifications.

High pressure steam, 1525 PSIA and 585°F, is generated by waste heat recovery in the CO shift and methanation units. This steam is superheated in a gas (CH_4) fired superheater and used for turbine drivers and the CO shift steam requirement. Process condensate, not used for cooling tower make-up water, boiler blowdown, and water treatment sludge are treated within the alternate product module. Solids are routed to the Lurgi base facility for disposal and the excess treated water (315 GPM) can be used in the Lurgi base facility to reduce import raw water or dumped. This alternate product design requires 266 GPM of raw water only during start-up. The water requirements for this facility can be met by utilizing the process generated condensate.



(1) For details, refer to Steam Balance Diagram, DWG NO. 517-SK-001 B.

(2) For details, refer to Water Balance Diagram, DWG NO. 517-SK-001 C.

(3) Raw water requirements for start-up only - 266 GPM.

(4) Losses include blowdown, drift and evaporation.

(5) Process solids are routed to LURGI MBG facility for disposal.

(6) Acid Gas is handled in CLAUS unit of LURGI MBG facility.

BDM | ANTHEM/KAISER corporation

NASA CONTRACT NAS8-33824

COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY

5.2.4.

SHM

MDD/JVF

ALTERNATE PRODUCT ANALYSIS
LURGI ADD-ON CASE II-A
METHANE PRODUCTION
MODULE BLOCK FLOW DIAGRAM

571-SK-001A | 0 | 8 3 80

TABLE 3-1. CASE II-A MATERIAL BALANCE - METHANE PRODUCTION

STREAM NUMBER			A-1	A-2	A-3	A-4	A-5	A-6	A-7	A-8
COMPONENT (COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)	SYMBOL	M.W.	LURGI FACILITY NET MBG PRODUCT	H.P. STEAM TO CO SHIFT	CO SHIFT FEED	AGR FEED	ACID GAS TO LURGI BASE FACILITY	METHANATION UNIT FEED	METHANE PRODUCT TO GAS DRYING	PRODUCT GAS OUT OF DRYING
HYDROGEN	H_2	2.016	14,865.1		14,865.1	15,577.0	-	15,514.7	436.9	436.9
NITROGEN	N_2	28.016	140.2		140.2	140.2		140.0	140.0	140.0
CARBON MONOXIDE	CO	28.01	5,431.2		5,431.2	4,720.3		4,720.3	4.7	4.7
CARBON DIOXIDE	CO_2	44.01	8,268.4		8,268.4	8,980.3	3.5	236.8	78.6	78.6
HYDROGEN SULFIDE	H_2S	34.076	1.8		1.8	1.8	1.8	-	-	-
METHANOL	CH_3OH	32.042	-		-	-		-	-	-
CARBONYL SULFIDE	COS	60.075	5.0		5.0	5.0	5.0	-	-	-
METHANE	CH_4	16.042	2,850.0		2,850.0	2,850.0	-	2,850.0	8,151.6	8,151.6
ETHANE	C_2H_6	30.068	129.4		129.4	129.4	-	129.4	-	-
ETHYLENE	C_2H_4	28.052	84.5		84.5	84.5	-	84.5	-	-
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR		31,775.5		31,775.5	32,488.5	10.3	23,675.7	8,811.8	8,811.8
WATER	H_2O	18.016	4.7	21,274.2	21,733.7	72.11	-	-	8.1	1.3
TOTAL WET	lb MOLES/HR		31,780.2	21,274.2	53,509.2	32,560.6	10.3	23,675.7	8,819.9	8,813.1
TOTAL	M POUNDS/HR		602,343	391,383	993,727	616,411	516	229,818	139,308	139,185
MOLECULAR WEIGHT			18.95	18.0	18.56	18.93	50.07	9.71	15.8	15.79
TEMPERATURE, °F			120	900	430	120	120	120	100	100
PRESSURE, PSIA			615	1,500	575	500	30	320	1,025	1,015
MSCFH			12.06	8.25	20.3	12.36	0.10	8.98	3.35	3.34
GPM			-	-	-	-	-	-	-	-

TABLE 3-1. CASE II-A MATERIAL BALANCE - METHANE PRODUCTION (cont'd)

STREAM NUMBER			A-9	A-10	A-11	A-12	A-13	A-14	A-15	A-16
COMPONENT (COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)	SYMBOL	M.W.	METHANE PRODUCT	FUEL GAS TO SUPERHEATER	CO_2 VENT TO ATMOSPHERE	PROCESS CONDENSATE & BLOWDOWN	SOLIDS & WATER TO LURGI FACILITY	TREATED WATER TO RAW WATER SUPPLY	BOILER FEED WATER MAKE-UP	CO SHIFT CONDENSATE TO COOLING TOWER
HYDROGEN	H_2	2.016	412.0	24.9	62.3					
NITROGEN	N_2	28.016	132.0	8.0	0.2					
CARBON MONOXIDE	CO	28.01	4.4	0.3	-					
CARBON DIOXIDE	CO_2	44.01	74.1	4.5	8,740.0					
HYDROGEN SULFIDE	H_2S	34.076	-	-	-					
METHANOL	CH_3OH	32.042	-	-						
CARBONYL SULFIDE	COS	60.075	-	-						
METHANE	CH_4	16.042	7,687.6	463.7						
ETHANE	C_2H_6	30.068	-	-						
ETHYLENE	C_2H_4	28.052	-	-						
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR		8,310.4	501.4	8,802.5					
WATER	H_2O	18.016	1.3	0	72.1					
TOTAL WET	lb MOLES/HR		8,310.7	501.4	8,874.6					
TOTAL	M POUNDS/HR		113,198	7,987	386,078					
MOLECULAR WEIGHT			15.79	15.79	43.50	-	-	-	-	-
TEMPERATURE, °F			100	100	120	100	100	100	250	90
PRESSURE, PSIA			1,015	1,015	30	115	65	65	30	65
MMSCFH			3.15	0.19	3.28	-	-	-	-	-
GPM			-	-	-	1,043	758	265	266	625

3.3.2 Lurgi Base to Methane/Methanol Co-Production - Case II-B

Case II-B was designated as the analysis of facilities to produce methane and methanol from Lurgi-derived MBG. This decision was made due to the characteristics of the MBG and the methanol synthesis unit. The Lurgi-derived MBG developed for this study contained 8.97 per cent methane and 0.68 per cent ethane/ethylene and has a H_2/CO ratio of 2.74. The methanol synthesis unit requires an H_2/CO ratio of 2.5 and, consequently, does not require a CO shift unit. Furthermore, the purge and off-gases from the methanol synthesis unit are suitable for conversion to methane and the methane and heavier hydrocarbons are unaffected by the methanol catalytic reaction. The required process units are shown on Drawing 571-SK-002A. The associated material balance is provided in Table 3-2.

Rather than feed the methane portion of the Lurgi MBG to a methanation unit to catalytically convert the methane to CO and H_2 and since TVA has an identified market for synthetic natural gas (SNG), it was decided to use the methanol synthesis off-gases to produce a marketable SNG.

Lurgi-derived MBG at 100°F and 615 PSIA are fed to a selective rectisol unit, which is used to remove CO_2 to 1.0 per cent of the exit gas and essentially all of the sulfur compounds. The sulfur-free synthesis gas is then fed to the methanol synthesis unit where 93.14 per cent of the CO and 47.76 per cent of the CO_2 are converted, in the presence of H_2, into methanol and water. The crude methanol product, 97.6 per cent methanol, is routed to storage and shipping units.

The purge and expansion gases from the methanol synthesis unit exit at 95°F and 20 PSIA with an H_2/CO ratio of 11.66. For the methanation reaction to proceed efficiently, another source of carbon is required to utilize the H_2 content of the gas. CO is preferred for this reaction, but there is a pure CO_2 stream vented to the atmosphere from the Rectisol unit.

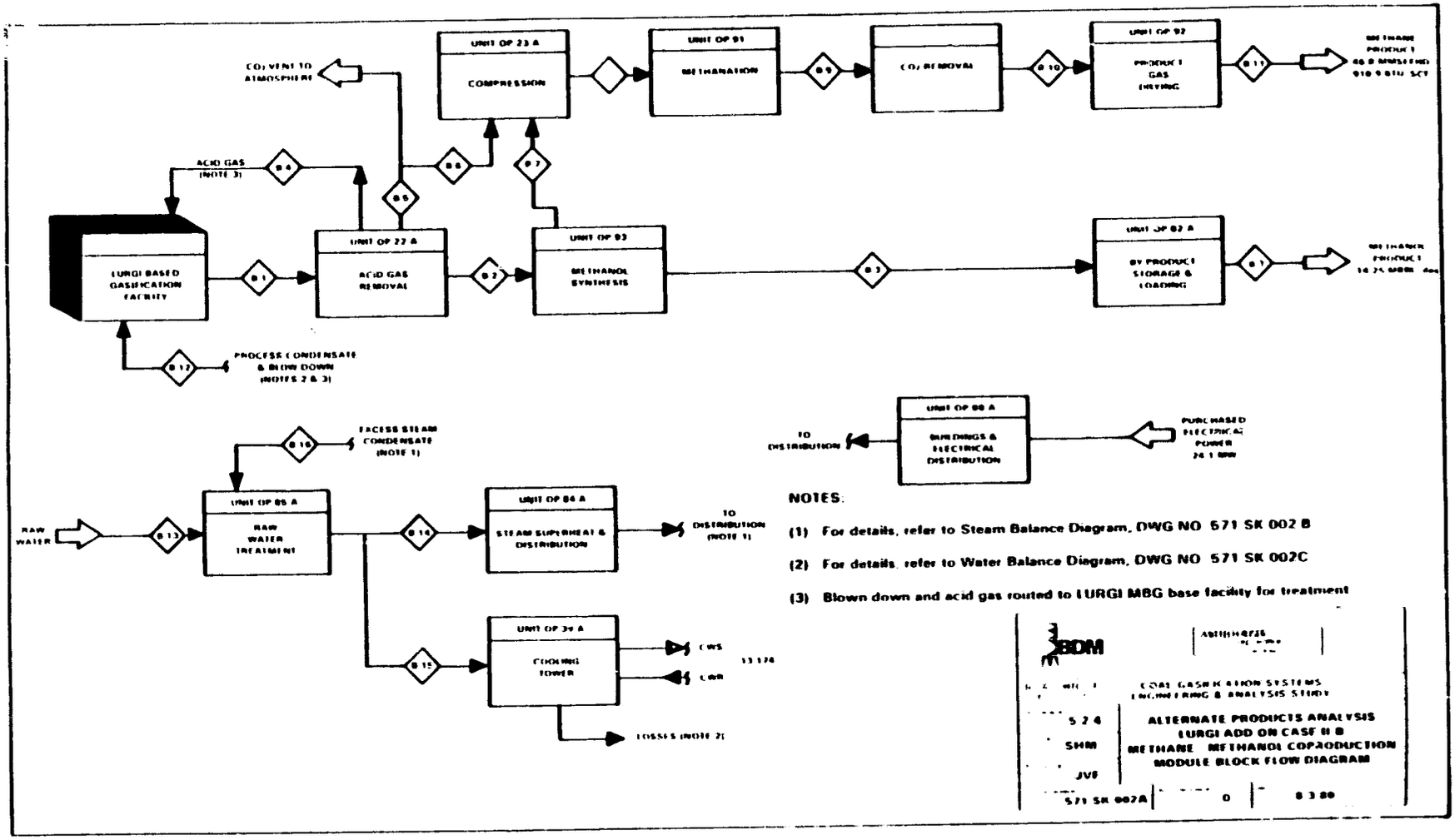

UNIT OP 23 A
COMPRESSION
UNIT OP 91
METHANATION
CO2 REMOVAL
UNIT OP 92
PRODUCT GAS DRYING
CO2 VENT TO ATMOSPHERE
ACID GAS (NOTE 3)
LURGI BASED GASIFICATION FACILITY
UNIT OP 22 A
ACID GAS REMOVAL
UNIT OP 93
METHANOL SYNTHESIS
UNIT OP 82 A
BY PRODUCT STORAGE & LOADING
PROCESS CONDENSATE & BLOW DOWN (NOTES 2 & 3)
UNIT OP 88 A
BUILDINGS & ELECTRICAL DISTRIBUTION
TO DISTRIBUTION
PURCHASED ELECTRICAL POWER
EXCESS STEAM CONDENSATE (NOTE 1)
RAW WATER
UNIT OP 85 A
RAW WATER TREATMENT
UNIT OP 84 A
STEAM SUPERHEAT & DISTRIBUTION
TO DISTRIBUTION (NOTE 1)
UNIT OP 39 A
COOLING TOWER
CWS
CWR
LOSSES (NOTE 2)
NOTES:
(1) For details, refer to Steam Balance Diagram, DWG NO 571 SK 002 B
(2) For details, refer to Water Balance Diagram, DWG NO 571 SK 002C
(3) Blown down and acid gas routed to LURGI MBG base facility for treatment
BDM
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
ALTERNATE PRODUCTS ANALYSIS
LURGI ADD ON CASE II B
METHANE METHANOL COPRODUCTION
MODULE BLOCK FLOW DIAGRAM
571 SK 002A
0
8 3 80

TABLE 3-2. CASE II-B MATERIAL BALANCE - METHANE/METHANOL PRODUCTION

STREAM NUMBER			B-1	B-2	B-3	B-4	B-5	B-6	B-7	B-8
COMPONENT	SYMBOL	M.W.	LURGI FACILITY NET MBG PRODUCT TO AGR	AGR OUTLET TO METHANOL SYNTHESIS	METHANOL PRODUCT	ACID GAS TO LURGI FACILITY	CO_2 VENT STREAM	CO_2 FEED GAS TO COMPRESSION	PURGE AND EXPANSION GAS TO COMPRESSION	METHANATION UNIT FEED
(COMPONENT QUANTITIES REPORTED IN LB MOLES/HR)										
HYDROGEN	H_2	2.016	14,865.0	14,805.5	-	-	59.5	7.8	4,345.0	4,352.8
NITROGEN	N_2	28.016	140.2	139.9	-	-	0.3	0.1	139.9	140.0
CARBON MONOXIDE	CO	28.01	5,431.2	5,431.2	-	-	-	-	372.6	372.6
CARBON DIOXIDE	CO_2	44.01	8,268.4	236.8	8.4	3.5	8,028.1	1,050.0	115.3	1,165.3
HYDROGEN SULFIDE	H_2S	34.076	1.8	-	-	1.8	-	-	-	-
METHANOL	CH_3OH	32.042	-	-	5,106.7	-	-		65.0	65.0
CARBONYL SULFIDE	COS	60.075	5.0	-	-	5.0	-		-	-
METHANE	CH_4	16.042	2,850.0	2,850.0	2.4				2,847.9	2,847.9
ETHANE	C_2H_6	30.068	129.4	129.4	-				129.4	129.4
ETHYLENE	C_2H_4	28.052	84.5	84.5	-				84.5	84.5
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	LB MOLES/HR		31,775.5	23,677.3	5,117.2	10.3	8,087.9	1,057.9	8,099.7	9,157.5
WATER	H_2O	18.016	4.7	-	113.1	-	4.2	1.6	-	0.6
TOTAL WET	LB MOLES/HR		31,780.2	23,677.3	5,230.8	10.3	8,092.6	1,058.5	8,099.7	9,158.4
TOTAL	M POUNDS/HR		602,243	248,298	164,032	516	353,530	46,240	82,223	122,399
MOLECULAR WEIGHT			18.95	10.49	31.36	50.07	43.69	43.68	10.15	
TEMPERATURE, °F			120	75	95	75	75	75	75	262
PRESSURE, PSIA			615	575	20	30	30	30	30	364
MMSCFH			12.06	8.99	-	-	3.07	0.4	3.07	3.48
GPM			-	-	416	-	-	-	-	-

TABLE 3-2. CASE II-B MATERIAL BALANCE - METHANE/METHANOL PRODUCTION (cont'd)

STREAM NUMBER			B-9	B-10	B-11	B-12	B-13	B-14	B-15	B-16
COMPONENT / STREAM NAME	SYMBOL	M.W.	METHANATION PRODUCT GAS TO CO_2 REMOVAL	METHANE PRODUCT TO GAS DRYING	METHANE PRODUCT	PROCESS CONDENSATE & BLOWDOWN	RAW WATER MAKE-UP	BOILER FEED WATER MAKE-UP	COOLING TOWER MAKE-UP	EXCESS STEAM CONDENSATE
(COMPONENT QUANTITIES REPORTED IN lb MOLES/HR)										
HYDROGEN	H_2	2.016	467.6	467.6	467.6					
NITROGEN	N_2	28.016	140.0	140.0	140.0					
CARBON MONOXIDE	CO	28.01	0.2	0.2	0.2					
CARBON DIOXIDE	CO_2	44.01	364.1	2.6	2.6					
HYDROGEN SULFIDE	H_2S	34.076	-	-	-					
METHANOL	CH_3OH	32.042	-	-						
CARBONYL SULFIDE	COS	60.075	-	-						
METHANE	CH_4	16.042	4,514.6	4,514.6	4,514.6					
ETHANE	C_2H_6	30.068								
ETHYLENE	C_2H_4	28.052								
PROPANE	C_3H_8	44.094								
PROPYLENE	C_3H_6	42.078								
ISO BUTANE	C_4H_{10}	58.12								
N BUTANE	C_4H_{10}	58.12								
BUTENES	C_4H_8	56.104								
ISO PENTANE	C_5H_{12}	72.146								
N PENTANE	C_5H_{12}	72.146								
PENTENES	C_5H_{10}	70.13								
GASOLINE	C_6^+									
TOTAL DRY	lb MOLES/HR		5,486.5	5,125.0	5,125.0					
WATER	H_2O	18.016	5.1	4.7	0.8					
TOTAL WET	lb MOLES/HR		5,491.6	5,125.7	5,125.8					
TOTAL	M POUNDS/HR		93,410	77,493	77,423	85,981	394,913	61,486	328,928	21,292
MOLECULAR WEIGHT			17.01	15.12	15.10	18.0	18.0	18.0	18.0	18.0
TEMPERATURE, °F			100	100	100	100	70	250	100	130
PRESSURE, PSIA			1,035	1,025	1,015	40	50	1,550	75	100
MMSCFH			2.08	1.95	1.95	-	-	-	-	-
GPM			-	-	-	172	790	123	658	43

4.0 DISCUSSION

This section of the report describes the methodology and detail of the process design of the two cases to produce methanol and/or methane. The designs presented herein are based on the net MBG product from one Lurgi gasification and converting the MBG into the alternate products. The full gas requirement for superheating steam has been taken from the methane product in both cases. This was done for three reasons; first, the higher heating value (HHV) of the methane product requires a lower flowrate of gas; second, the use of MBG would result in lower alternate product yield; and third, there was no appreciable cost saving on the units based on reduced through-put.

The off-site portions were designed to be self-sufficient in steam production and water treat. The acid gas from the Rectisol unit and certain blowdown streams were treated in the Lurgi base facility because: (1) it was not cost effective or practical to design units for such small flows; and (2) the flowrates would not seriously impact the units in the base facility.

4.1 Case II-A: Methane Production

Process Units

The production of methane from synthesis gas requires adjustment of the H_2/CO ratio to approximately 3.0-3.2. However, Lurgi-derived synthesis contains an appreciable amount of ethane and ethylene that cracks during the methanation reaction. For this reason, the H_2/CO ratio was set at 3.3 to provide excess H_2 to saturate the products of the cracking reaction.

The CO shift unit for this case was selected from a Lurgi design prepared by C. F. Braun[3]. The design conditions for the shift unit used a steam to dry gas ratio of 0.65 WT, and geared to adjust composition of the

synthesis gas to a H_2/CO ratio of 3.3. Since the MBG gas had a starting H_2/CO ratio of 2.74, only 711.5 moles of CO are shifted to CO_2.

The acid gas removal unit chosen was a selective Rectisol from a design prepared by Continental Oil[4]. The selective Rectisol unit was selected to meet the low sulfur level requirement imposed by the methanation catalyst. The Rectisol unit was placed after the shift unit so as to remove CO_2 and sulfur constituents at one time. The design of this unit results in all H_2S and COS removed to 1 PPMV or less, 0.4% H_2 loss, 0.19% N_2 loss, and CO_2 removed to only 10% of the outlet gas.

The CO_2 stream is routed to the atmosphere and the sulfur constituents are routed to the sulfur recovery unit in the Lurgi base facility. This acid gas stream is 38 per cent sulfur and is only 10.3 lb moles/hr. This small stream can be handled in the base plant facility.

The methanation unit selected for this study was a hot recycle case from a C. F. Braun work[3]. This design includes a trim methanation reactor and a compressor to boost the gas pressure to the required transmission pressure. Methane formed from Lurgi gas is from two sets of reactions. The first type is the synthesis gas reactor, converting CO and CO_2, in the presence of H_2, into methane and water.

$$CO + 3H_2 \rightleftharpoons CH_4 + H_2O$$

$$CO_2 + 4H_2 \rightleftharpoons CH_4 + 2H_2O$$

The second reaction type is the thermal cracking (decomposition) of C_2 hydrocarbons according to the following reactions.

$$C_2H_4 + 2H_2 \longrightarrow 2CH_4$$

$$C_2H_6 + H_2 \longrightarrow 2CH_4$$

The total moles of CH_4 formed was 5301.6 lb mole/hr.

The last process unit required in this study was the SNG drying unit. This unit uses a triethyleneglycol solution to dry the gas by absorbing water to result in a gas with 7 lbs H_2O per million cubic feet of gas.

Utility Units

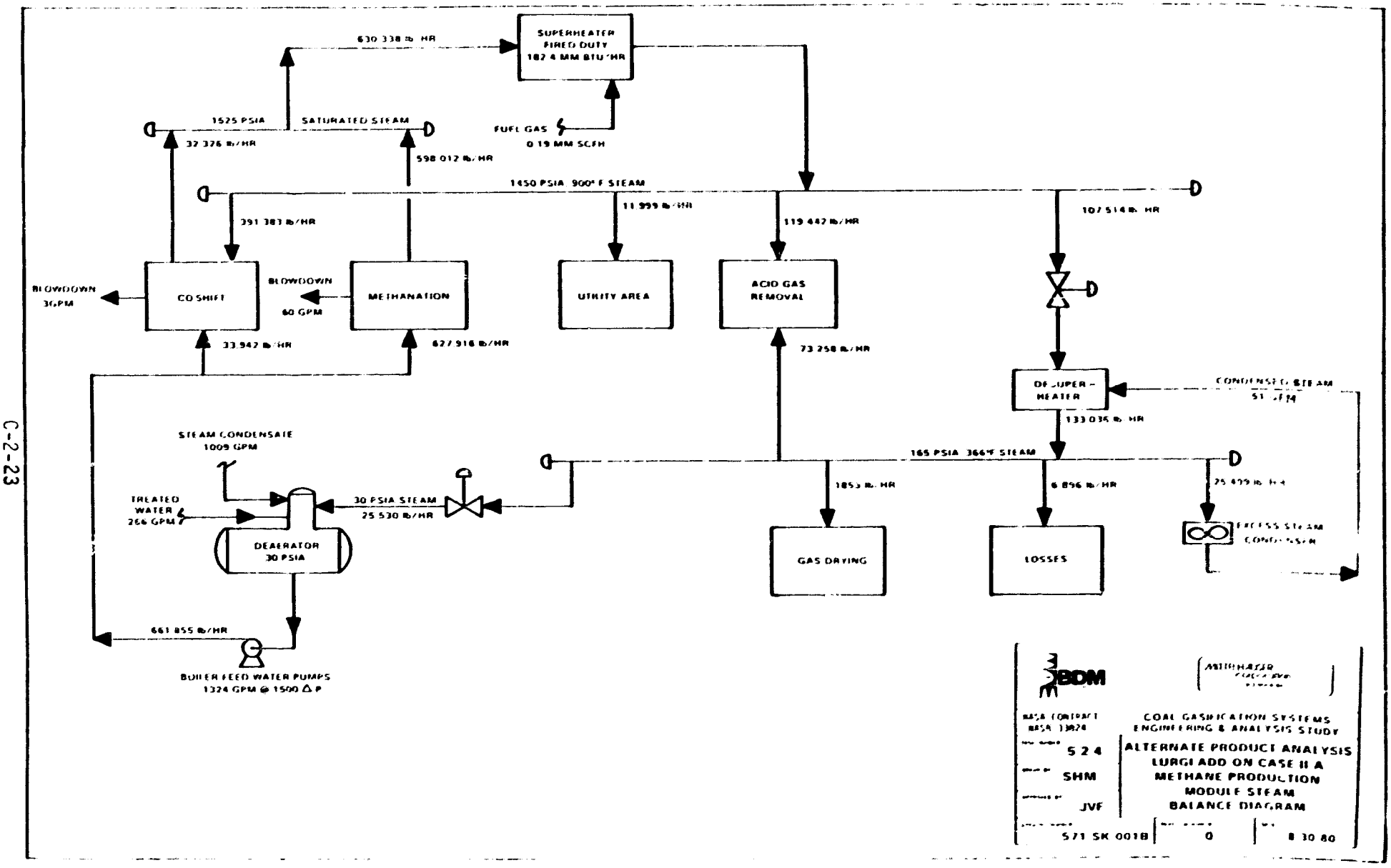
The utility units required to support the methanation facilities were designed with minimum duplication of the base facility.

High pressure (HP) steam is generated at 1525 PSIA, saturated by waste heat recovery in the CO shift and methanation units. This steam is superheated in an SNG fired heater and distributed at 1450 PSIA and 900°F for use in turbine drivers. The superheater uses 501 lb moles/hr (.19 MMSCFH) of SNG at a fired duty of 182.4 MM Btu/hr. LP steam requirements are met by desuperheating 107,514 lbs/hr of HP steam to 165 PSIA and 366°F. Excess LP steam is condensed and used as desuperheating water. The boiler feed water pumps supply 1324 GPM of BFW for steam generation. Drawing 571-SK-001B depicts the methane alternate product module steam balance.

Process condensate from the CO shift unit is used as cooling tower make-up water. It is felt that the stream is not sufficiently contaminated to cause environmental problems. The cooling tower has been sized to provide 13,497 GPM of cooling water to process users.

Excess condensate from CO shift (350 GPM), methanation process condensate (181 GPM), boiler blowdown (63 GPM) and water treating sludge (19 GPM) are treated in a solids treatment facility prior to reuse or discharge.

The water treatment facilities were designed to use 285 GPM of treated water from solid treatment for use as boiler feed water make-up. It


630 338 lb/HR
SUPERHEATER
FIRED DUTY
182 4 MM BTU/HR
1525 PSIA
SATURATED STEAM
32 376 lb/HR
598 012 lb/HR
FUEL GAS
0 19 MM SCFH
1450 PSIA 900°F STEAM
391 387 lb/HR
11 999 lb/HR
119 442 lb/HR
107 514 lb/HR
BLOWDOWN
3GPM
CO SHIFT
BLOWDOWN
60 GPM
METHANATION
UTILITY AREA
ACID GAS
REMOVAL
33 942 lb/HR
627 916 lb/HR
73 258 lb/HR
DE-SUPER-
HEATER
CONDENSED STEAM
51 GPM
133 035 lb/HR
STEAM CONDENSATE
1009 GPM
165 PSIA 366°F STEAM
TREATED
WATER
266 GPM
30 PSIA STEAM
25 530 lb/HR
DEAERATOR
30 PSIA
185 lb/HR
6 896 lb/HR
25 479 lb/HR
PROCESS STEAM
CONDENSATE
GAS DRYING
LOSSES
661 855 lb/HR
BOILER FEED WATER PUMPS
1324 GPM @ 1500 Δ P
BECHTEL
NASA CONTRACT
NASA 13824
COAL GASIFICATION SYSTEMS
ENGINEERING & ANALYSIS STUDY
524
SHM
JVF
ALTERNATE PRODUCT ANALYSIS
LURGI ADD ON CASE II A
METHANE PRODUCTION
MODULE STEAM
BALANCE DIAGRAM
571 SK 0018
0
8 30 80

was assumed for this study that 80 per cent of the steam condensate would be returned for BFW. There is a demand for 266 GPM of raw water import for start-up only.

The solids from solids treating are routed to the Lurgi base facility for disposal. The excess treated water (315 GPM) can be routed to the Lurgi base facility to reduce raw water import or discarded to the Tennessee River. Drawing 571-SK-001C depicts the alternate product module water balance.

Utility Summary

Table 4-1 summarizes the utility requirements of the methane alternate product module process and utility units.

4.2 Case II-B: Methane-Methanol Co-Production

This case was originally intended to be a study of the production of methanol. However, due to the fact that the Lurgi MBG contained 8.96 MOL per cent CH_4 and the methanol synthesis off-gases are suitable for methane synthesis, it was decided to provide the facilities to produce methane as well as methanol. The only other decision would be to use the methanol synthesis purge-gas in the fuel system or convert it to synthesis gas. This would have been an ineffective use of the methane content of the MBG.

Process Units

The processing sequence required for producing methanol requires no CO shift unit to adjust the H_2/CO ratio. The methanol synthesis unit selected requires an H_2/CO ratio of 2.5 and the synthesis gas is already at a 2.74 H_2/CO ratio. The methanol synthesis reaction converts CO and CO_2, plus H_2 into methanol and water. The assumption was made that the unit configuration leaves little if any gases in the liquid phase and that the

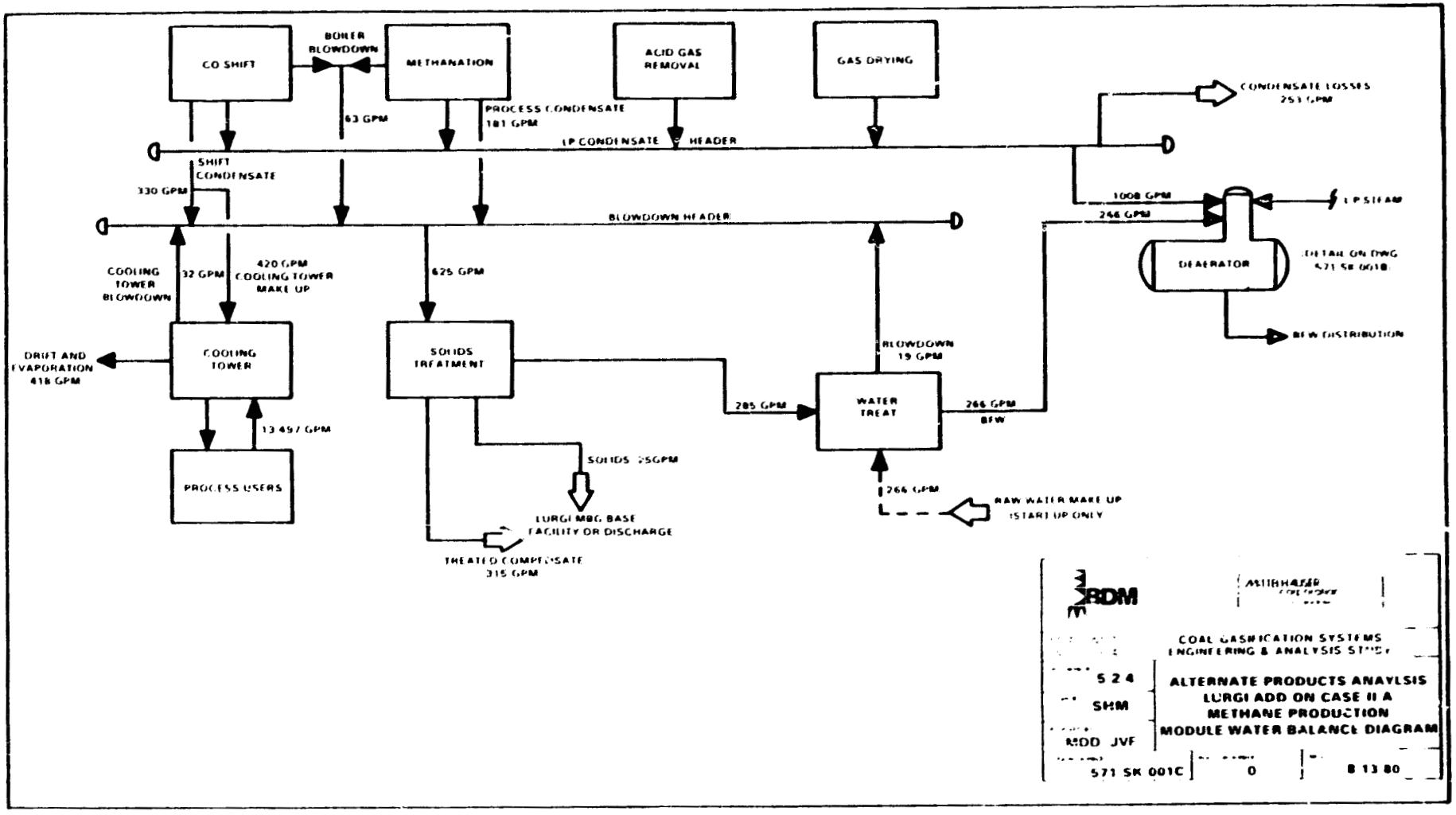

CO SHIFT
BOILER BLOWDOWN
METHANATION
ACID GAS REMOVAL
GAS DRYING
CONDENSATE LOSSES 253 GPM
63 GPM
PROCESS CONDENSATE 181 GPM
LP CONDENSATE HEADER
SHIFT CONDENSATE
330 GPM
1008 GPM
LP STEAM
246 GPM
BLOWDOWN HEADER
DEAERATOR
DETAIL ON DWG 571 SK 001B
COOLING TOWER BLOWDOWN
32 GPM
420 GPM COOLING TOWER MAKE UP
625 GPM
BFW DISTRIBUTION
BLOWDOWN 19 GPM
DRIFT AND EVAPORATION 418 GPM
COOLING TOWER
SOLIDS TREATMENT
285 GPM
WATER TREAT
266 GPM BFW
13 497 GPM
PROCESS USERS
SOLIDS 25GPM
266 GPM
RAW WATER MAKE UP (START UP ONLY)
LURGI MBG BASE FACILITY OR DISCHARGE
TREATED CONDENSATE 315 GPM
BDM
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5 2 4
SHM
MDD JVF
ALTERNATE PRODUCTS ANAYLSIS
LURGI ADD ON CASE II A
METHANE PRODUCTION
MODULE WATER BALANCE DIAGRAM
571 SK 001C
0
8 13 80

TABLE 4-1. UTILITY SUMMARY - METHANE ALTERNATE PRODUCT MODULE

UNIT OP NUMBER	UNIT/CONSUMER	ELECTRICAL POWER KW	COOLING WATER GPM	DRIVER HP BHP	HEAT TO EXPORT 10^6 BTU/HR	IMPORT LP STEAM lb/HR	IMPORT HP STEAM lb/HR	EXPORT HP STEAM lb/HR	PROCESS CONDENSATE GPM
90	CO SHIFT	49.8	356	-	24.9	-	391,383	32,326	780
22-A	ACID GAS REMOVAL	3,788.9	12,907	18,375.6	-	73,258	119,442	-	-
91	METHANATION	12,074.6 109.7	217	16,185.8	460.9	-	-	598,012	181
92	GAS DRYING	151.0	17	-	-	1,853	-	-	-
	SUBTOTAL	16,174	13,497			75,111	510,825	630,338	
39-A	COOLING H_2O SUPPLY	*	*	1,255.6	-	-	8,159	-	
85-A	WATER TREATING	*	*	591.0	-	25,530	3,840	-	
	SUBTOTALS						11,999		
	TOTALS	20,217.5**	13,497			100,641	522,824	630,338	

* Included in 1.25 factor
** Total electrical requirement multiplied by 1.25 to account for motor losses, lighting, etc.

water and methanol are in solution. The crude methanol product is sent directly to storage for shipment.

The purge and off-gas streams leave the methanol unit at 20 PSIA and rich in CH_4 and H_2. These streams are suitable for conversion into methane with some alteration. The H_2/CO ratio is excessively high at 11.7. The high hydrogen content would result in a methane produce with an unacceptably low HHV.

One method of improving the quality of the methane product is to add carbon species such as CO or CO_2 to effectively use the hydrogen in the methanation. CO is more amenable to the reaction; but, in this study, a relatively pure stream of CO_2 is available from the rectisol CO_2 vent stream. The CO tends to react at 99.9% conversion while CO_2 is reacted at 68.8% conversion. This lower conversion for CO_2 results in an increased CO_2 content of the methane product. This additional CO_2 is easily removed in a MEA wash or CO_2 removal unit.

The methanation reaction takes place at high pressure, so the methanol synthesis purge gases and CO_2 vent gas are mixed at near atmospheric pressure and compressed to the necessary reaction pressure.

The methanation unit for this case is the same as in case II-B. The methanation reaction products are routed to a CO_2 removal unit and then to a TEG gas drying unit to provide a methane product of acceptable quality.

Utility Units

The utility units required to support the production of methane and methanol were designed with minimal duplication of the units in the Lurgi base facility.

High pressure (HP) steam is generated at 1525 PSIA, saturated by waste heat recovery in the methanol synthesis and methanation units. This steam is superheated in an SNG fired heater and distributed at 1450 PSIA and 900°F for use in turbine drivers. The superheater uses 339 lb moles/hr (.13 MMSCFH) of SNG at a fired duty of 118.1 MM Btu/hr. LP steam requirements are met by desuperheating 130,694 lbs/hr of HP steam to 165 PSIA and 366°F. Excess LP steam is condensed and used to displace raw water in water treating. The boiler feed water pumps supply 866 GPM of BFW for steam generation. Drawing 571-SK-002B depicts the methane/methanol alternate product module steam balance.

The cooling tower has been sized to provide 13,174 GPM of cooling water to process users.

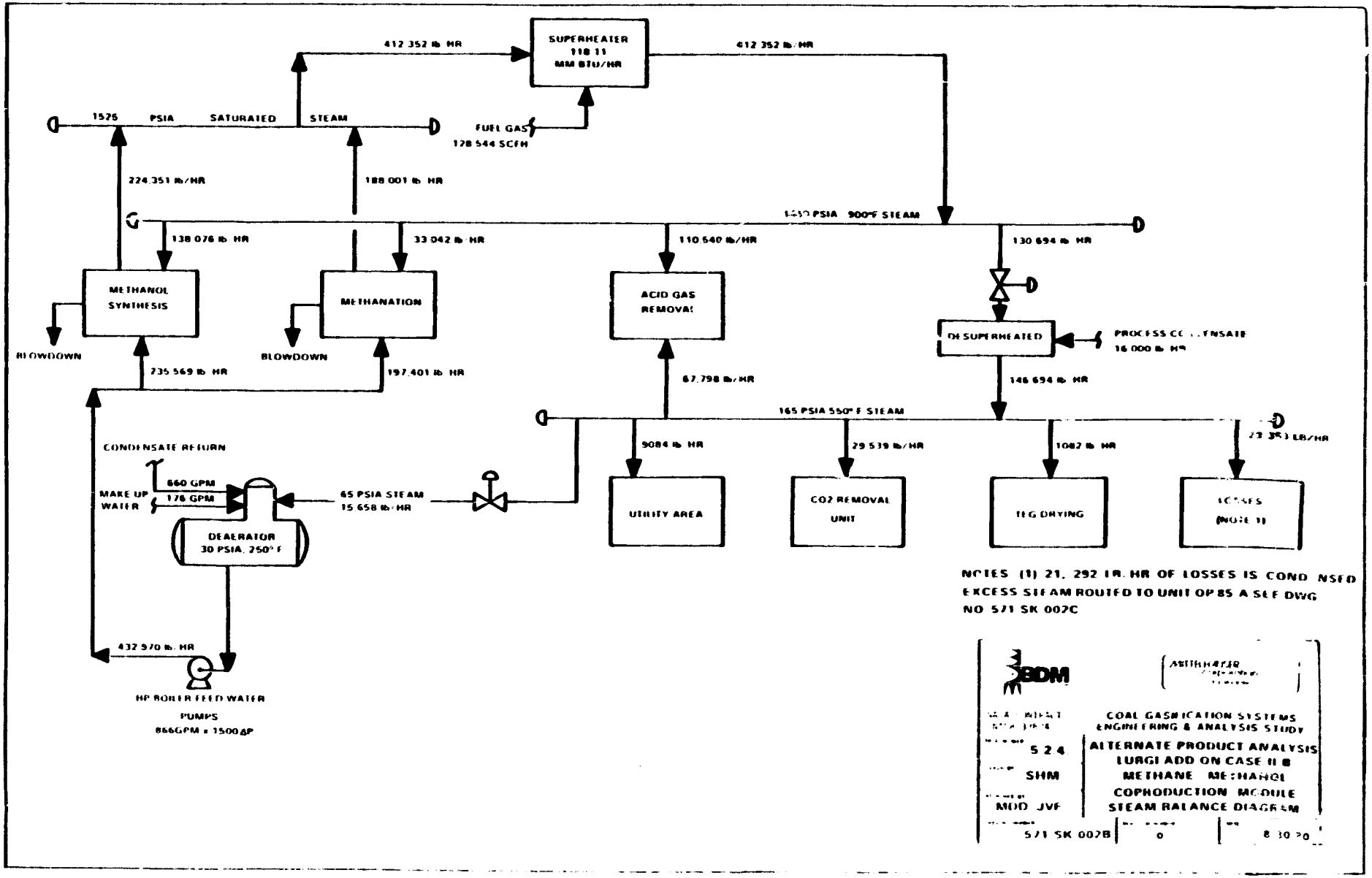
Process condensate from methanation (74 GPM) and methanol synthesis (5 GPM), boiler blowdown (41 GPM) and water treating sludge (52 GPM) are routed to the Lurgi base facility for treatment.

The water treatment facilities were designed to supply 439 GPM cooling tower make-up water and 123 GPM of boiler feed water make-up. It was assumed for this study that 80 per cent of the steam condensate would be returned for BFW. There is a demand for 571 GPM of raw water import.

Drawing 571-SK-002C depicts the alternate product module water balance.

Utility Summary

Table 4-2 summarizes the utility requirements of the methane alternate product module process and utility units.



NOTES (1) 21, 292 LB HR OF LOSSES IS CONDENSED EXCESS STEAM ROUTED TO UNIT OP 85 A SEE DWG NO 571 SK 002C

BDM

COAL GASIFICATION SYSTEMS
ENGINEERING & ANALYSIS STUDY

524
SHM
MOD JVF

ALTERNATE PRODUCT ANALYSIS
LURGI ADD ON CASE II B
METHANE METHANOL
COPRODUCTION MODULE
STEAM BALANCE DIAGRAM

571 SK 002B | 0 | 8 30 80

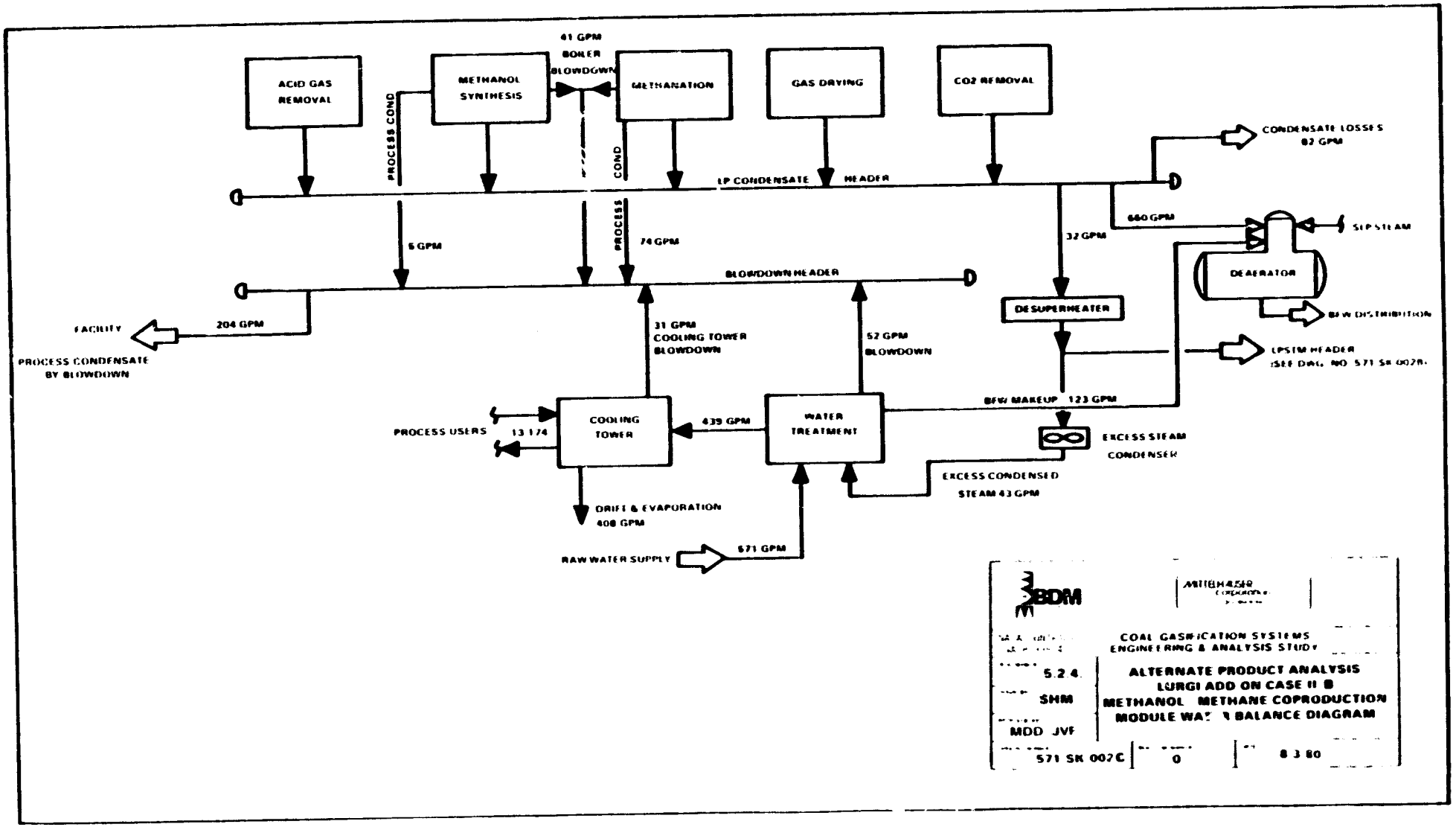

ACID GAS REMOVAL
METHANOL SYNTHESIS
41 GPM BOILER BLOWDOWN
METHANATION
GAS DRYING
CO2 REMOVAL
PROCESS COND
COND
CONDENSATE LOSSES 82 GPM
LP CONDENSATE HEADER
660 GPM
SLP STEAM
32 GPM
5 GPM
PROCESS
74 GPM
BLOWDOWN HEADER
DEAERATOR
DESUPERHEATER
BFW DISTRIBUTION
FACILITY
204 GPM
PROCESS CONDENSATE BY BLOWDOWN
31 GPM COOLING TOWER BLOWDOWN
52 GPM BLOWDOWN
LPSTM HEADER
BFW MAKEUP 123 GPM
PROCESS USERS 13 174
COOLING TOWER
439 GPM
WATER TREATMENT
EXCESS STEAM CONDENSER
EXCESS CONDENSED STEAM 43 GPM
DRIFT & EVAPORATION 408 GPM
RAW WATER SUPPLY
571 GPM
BDM
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
SHM
MDD JVF
ALTERNATE PRODUCT ANALYSIS
LURGI ADD ON CASE II B
METHANOL METHANE COPRODUCTION
MODULE WATER BALANCE DIAGRAM
571 SK 007C
0
8 3 80

TABLE 4-2. UTILITY SUMMARY - METHANE/METHANOL ALTERNATE PRODUCT MODULE

UNIT OP NUMBER	UNIT/CONSUMER	ELECTRICAL POWER KW	COOLING WATER GPM	DRIVER HP BHP	HEAT TO EXPORT 10^6 BTU/HR	LP STEAM IMPORT lb/HR	HP STEAM IMPORT lb/HR	HP STEAM EXPORT lb/HR	PROCESS CONDENSATE GPM
22-A	ACID GAS REMOVAL	3,506.5	11,944	17,006.2	-	67,798	110,540	-	-
93	MEOH SYNTHESIS	-	844	20,934.2	172.93	-	136,078	224,351	5*
91	METHANATION	41.2	68	5,088	144.91	-	33,042	188,001	74
-	CO_2 REMOVAL	327.9	268	-	-	29,539	-	-	-
92	TEG DRYING	88.1	10	-	-	1,082	-	-	1
82-A	BY-PRODUCT STORAGE & LOADING	149.1	-	-	-	-	-	-	-
23-A	COMPRESSION	15,176.6	-	20,221	-	-	-	-	-
	SUBTOTAL	19,289.5	13,174			98,414	279,660	412,352	
	WATER TREATMENT	**							
	STEAM SYSTEM	**				5,337			
	COOLING TOWER					3,747			
	SUBTOTALS					9,084			
	TOTAL	24,106.9***	13,174			107,498	279,660	412,352	80

* Guess
** Included in 1.25 factor
***Electric power consumption multiplied by 1.25 to account for motor losses, lighting, etc.

5.0 REFERENCES

1. Facility Definition Designs - Lurg, BGC Lurgi, and B&W, Volume I, BDM/H-80-590-TR, September 3, 1980

2. Alternate Product Analysis - Task 5.1.4, BDM/H-80-481-TR, July 24, 1980

3. FE-2240-31, Factored Estimates for Eastern Coal Commercial Concepts, Interim Report, C. F. Braun, 1978

4. FE-2542-10, The Pipeline Demonstration Plant, Vol. 2, Process and Project Engineering Design, Continental Oil Company, 1978

5. FE-2247-13, Research Guidance Studies to Assess Gasoline from Coal by Methanol-to-Gasoline and Sasol-Type Fischer-Tropsch Technologies, Final Report, Mobil Research and Development Company, 1978

APPENDIX C-3
MBG/SNG FACILITIES:
KOPPERS-TOTZEK AND KOPPERS-TOTZEK/TEXACO

TABLE OF CONTENTS

TABLE OF CONTENTS (Concluded)

LIST OF TABLES

LIST OF TABLES (Concluded)

LIST OF DRAWINGS

1.0 INTRODUCTION

This report covers the work done to develop two designs for plants to manufacture substitute natural gas (SNG) from coal. This work is to be used by NASA in their support of the TVA-20,000 TPD* coal gasification project. Each design covers a complete grassroots facility; the plant inputs are coal, raw water, and electric power. The plant outputs are SNG, medium-BTU gas (MBG), and prilled sulfur.

Each of the plant designs was based on a different coal gasification technology. One of the plants is based on Koppers-Totzek (K-T) technology; the other on a combination of K-T and Texaco technology. Each plant was divided into four parallel process modules, each capable of handling 5,000 TPD of coal.

The designs for the production of SNG from MBG are intended as an extension of a previous contract subtask in which preliminary estimates of alternate product cost were made for K-T and Texaco facilities. As such, they used the Reference Facility Designs for these two technologies as a starting point. These Reference Facility Designs are described in detail in a separate report.[1,2]

The overall design approach utilized parallel modules of a capacity sufficient to process the MBG from a single gasification module. Each module was divided into process and utility systems.

For each process system, available technologies were identified and comparative tradeoff studies performed to select the preferred technologies for design. Each system's design was based on published data, engineering experience and judgement, and design calculations specific to this process.

Cost estimates were based on conceptual level design estimating procedures as described in an earlier report.[3] Cost results are presented in Volume II and Volume III.

*TPD indicates Tons Per Day.

2.0 SUMMARY

The two MBG Upgrading Plant designs reported herein each comprise two parallel modules of processing systems designed to convert medium-BTU gas (300 BTU/SCF) into SNG (920 BTU/SCF). One of the designs is for a facility consisting of four MBG modules based on Koppers-Totzek technology. The other design is for a facility consisting of one Koppers-Totzek and three Texaco modules. Each of the MBG Upgrading Plants is designed according to the TVA Facility Design Basis[6] and the guidelines reported below.

- The facility's total capacity, based on a nominal 20,000 TPD of feed coal, is
  - up to 100% MBG, or
  - 50% MBG and 50% SNG.
- The first two facility modules must be designed to produce 100% MBG, 100% SNG, or a mixture of both.
- Any of the four facility modules must be capable of feeding the MBG Upgrading Plant.
- The MBG Upgrading Plant shall be integrated with the remainder of the Coal Gasification Facility, rather than being designed as an add-on plant.

The two parallel modules in the MBG Upgrading Plant consist of a CO Shift Conversion System, an Acid Gas Removal System, and a Methanation System. Table 2.1 summarizes the results of the design and cost work presented herein. Table 2.2 describes the quality specifications for the product SNG.

Table 2.1

Study Results

	Design #1 Based On 2 K-T Modules To SNG	Design #2 Based on 1 Texaco and 1 K-T Module Each To SNG
MBG to Upgrading Plant SCFD BTU/Day	529.92×10^6 160.84×10^9	536.76×10^6 162.71×10^9
SNG from Upgrading Plant SCFD BTU/Day	114.20×10^6 106.20×10^9	122.20×10^6 114.23×10^9
MBG from Modules 3&4 SCFD BTU/Day	459.60×10^6 140.18×10^9	538.80×10^6 156.63×10^9
Capital Requirement 1980 Dollars - Total Facility	$2,579.00 \times 10^6$	$2,407.00 \times 10^6$
Annual Operating and Maintenance Costs, 1980 Dollars - Total Facility	389.00×10^6	339.00×10^6
Annual Coal, Catalyst and Chemicals Costs, 1980 Dollars - Total Facility	184.00×10^6	184.00×10^6
Purchased Electricity, KWH per Year - Total Facility	3.6×10^9	2.1×10^9
Raw Water Consumption, Gallons per Year - Total Facility	9.7×10^9	4.5×10^9
Operating Staff - Total Facility	400	388
Product Price 1980 $/MMBtu	8.02	6.49

Table 2.2

SNG Quality Criteria

	Preferable Value	Objectionable Value
Lifting Index	Under 1.0	Above 1.06
Flash-Back Index	Under 1.18	Above 1.20
Yellow Tip Index	Above 1.00	Under 0.80

Source: Reference 7

3.0 DESIGN BASIS AND APPROACH

3.1 Background

In order to broaden the potential markets for products from their 20,000 TPD Coal Gasification Facility, TVA is considering the manufacture of alternate products within the facility, using the MBG produced in the facility as a feedstock. Among the alternate products under consideration are SNG, methanol, gasoline and hydrogen. Earlier tasks consisted of preliminary investigation of the production costs of each product using K-T and Texaco, and Lurgi gasification technologies. The results of these tasks have previously been reported.[4,5]

The studies, referred to above, identified SNG as a product with attractive costs. Since TVA had identified a market for SNG from the facility, a design for facilities to produce this product, based on the already completed Reference Facility Designs, was desirable to provide data to TVA concerning requirements and costs for producing SNG from MBG, at a level of detail comparable to those of the Reference Facility Designs. This design work was therefore performed, in lieu of a previously planned design effort to synthesize a Reference Facility Design for two gasification processes based on both NASA and other TVA contractors' conceptual designs.[9]

3.2 Design Basis

The design basis for the work done in this study began with the over-all TVA Facility Design Basis.[6] The major points are as follows:

- The overall facility feed is Kentucky No. 9 bituminous coal, with design provisions for handling up to five percent North Alabama coal. The total capacity of the facility is 20,000 TPD to the gasifiers.
- The facility design is structured as four parallel modules, each with a nominal design capacity of approximately 5000 TPD of coal to the gasifier.

- Sulfur is recovered for sale in prilled form. Ash is disposed on-site.
- MBG from the facility shall have a maximum sulfur content of 200 ppmv and a minimum higher heating value of 285 BTU/SCF.

As a result of earlier process tradeoff studies,[8] the following additional bases were incorporated into the designs for the MBG Upgrading Plants:

- The facility is designed for a zero liquid discharge to the nearby Tennessee River or to underground water supplies.
- Onsite steam shall not be generated by the combustion of coal in boilers.
- Electricity from the TVA grid is used to supply all plant prime mover power requirements above that which can be supplied by steam which is
  - generated from process waste heat
  - superheated in an MBG fired superheater.

In their planning for the 20,000 TPD Coal Gasification Project, TVA is considering two scenarios for plant design and construction. The first of these envisions the employment of four identical modules utilizing the Koppers-Totzek gasification process. The second scenario uses the Koppers-Totzek process in the first module, then switches to the Texaco process for the other three. TVA desires that the facility have the capability of producing between 50% and 100% of its total output as MBG, and between 0 and 50% as SNG. The first two modules, in addition, must together be capable of producing either 100% SNG, 100% MBG, or any mixture of both. When all four MBG modules are complete, any module must be capable of feeding the MBG Upgrading Plant.

3.3 Approach

Table 3.1 contains a design task that began with the already-completed Reference Facility Designs.[1,2] These designs were examined for integration possibilities in the light of other previous analyses of SNG as an alternate product from K-T and Texaco gasification.[4] The examination revealed that the major candidate for system integration was System 4, Acid Gas Removal. The manufacture of SNG requires a methanation catalyst which has little or no sulfur tolerance, therefore essentially complete removal of all sulfur compounds is required upstream of the Methanation System. For this reason, the Acid Gas Removal Systems in the Reference Facility Designs, which had been designed for a 200 ppmv treated gas sulfur concentration, were replaced by systems designed for a maximum treated gas sulfur concentration of 1 ppmv. The remainder of the sulfur removal was designed to be done in the Methanation System. Because of the requirement that each module be capable of feeding the MBG Upgrading Plant, all of the Acid Gas Removal System in the facility were replaced. Other than the Steam System, integration of other systems between the MBG Facility and the Upgrading Plant was purposely minimized to allow for the MBG/SNG flexibility desired by TVA.

Tradeoff Studies

Process tradeoff studies had been performed earlier to establish process choices for the following plant systems:

- CO Shift
- Acid Gas Removal
- Methanation
- Raw Gas Compression

Briefings have been presented to NASA discussing the results of these process tradeoffs.[8] Tables 3.2 through 3.5 give the results of the trade-off studies as reported at these briefings.

Some modifications to the conclusions presented at the briefings were made; these concerned CO shift and methanation. For CO shift, a

Table 3.1

Summary of Design Approach

1. Integration possibilities.

2. Tradeoff studies
   a. Select high-temperature Co-Mo shift catalyst
   b. Select Rectisol for AGR System
   c. Select Cold Recycle Methanation
   d. Place K-T Raw Gas Compression ahead of CO shift.

3. Do energy and material balances for CO shift and methanation systems. Use process simulation to establish heat exchanger configurations and duties.

4. Size equipment in CO shift and methanation systems.

5. Define Rectisol system material balance and interfaces.

6. Prepare process flow diagrams for CO shift and methanation systems.

7. Define utility and operating requirements for MBG Upgrading Plant.

8. Define incremental capital and operating costs for MBG Facility attributable to the MBG Upgrading Plant.

9. Define incremental capital and operating costs for MBG Upgrading Plant.

Fluor design[10] similar to that employed by Braun[11] was used as a starting point due to team members' familiarity of the process with its design basis. No significant difference between this design and the Braun design was observed. Both designs employ COS hydrolysis on the bypass gas, and a high-temperature cobalt-molybdate shift conversion catalyst.

Although the methanation scheme recommended in the process tradeoff studies was a hot recycle methanation scheme, the cold recycle scheme (see Table 3.4) is somewhat simpler and is approximately equal in capital cost. The virtue of the hot recycle scheme is that it results in a higher export of high pressure steam than the cold recycle scheme. However, this steam must be superheated in an MBG-fired superheater, which may reduce its value as an exportable product. A cold-recycle scheme based on a previous Fluor study[10] was selected as the basis for the design work in this study because;(1) it had been performed by team members, and; (2) it provided specific design and sizing data that could be used to specify the individual equipment items in the unit.

Because of the stringency of the sulfur removal requirement, the Rectisol process was selected over SELEXOL, even though COS hydrolysis was employed in the shift conversion system. This extra measure of conservation will not have a major impact on the overall plant cost. The Acid Gas Removal design was based on an in-house Rectisol design done for a Lurgi SNG plant. It was adjusted to account for the non-use of CO_2 as a pressurizing gas in the K-T and Texaco gasification process (no lock hoppers are used in these processes) and for the elimination of a naphtha prescrubbing section (there is no naphtha in the MBG from these two processes). The design is a selective configuration which produces an acid gas rich enough in H_2S to use as the Claus plant feed, and a waste CO_2 stream suitable for venting. It was assumed for this study that there would be no impact on System 5, Sulfur Recovery and Tail Gas Cleanup, because of the increase in stringency of sulfur removal from 200 ppmv to 1 ppmv in the treated gas.

Table 3.2

Shift Conversion

Evaluation & Decision

- EVALUATION
  - IRON-CHROME CATALYSTS HAVE APPROXIMATELY THE SAME COST AS COLBALT-MOLY; PARTIAL DEACTIVATION OF IRON-CHROME OFFSETS ITS LOWER COST PER POUND.
  - COBALT-MOLY CATALYSTS OFFER THE POSSIBILITY OF COS HYDROLYSIS WHICH CAN LOWER AGR COSTS.
  - DESIGN DONE BY TEAM MEMBERS USING Co-Mo IS AVAILABLE FOR THIS STUDY (BRAUN FE-2240-31).
- DECISION
  - USE COBALT-MOLY CATALYST IN BRAUN DESIGN CONFIGURATION FOR SHIFT CONVERSION.

Table 3.3

Acid Gas Removal Evaluation & Decision

- PREVIOUS TEAM STUDIES (FE-2240-49) HAVE SHOWN THAT
  - ATTRACTIVENESS OF PROCESSES VARIES WITH PARTIAL PRESSURE OF ACID GAS IN FEED.
  - AT PRESSURES NEAR THOSE OF THIS STUDY (600 PSIG) PREFERENCE IS BENFIELD, SELEXOL, RECTISOL.
  - COMPRESSION OF GAS PRIOR TO AGR IS ATTRACTIVE. EXTRA COST OF COMPRESSION IS MORE THAN OFFSET BY REDUCTIONS IN COST OF PHYSICAL SOLVENT.
  - FOR SULFUR REMOVAL FROM HIGH-SULFUR COAL GAS, SELECTIVE AGR IS PREFERRED.
  - BENFIELD AND SELEXOL HAVE PROBLEMS WHICH MAY LIMIT THEIR ABILITY TO GET TO <1 PPMV TOTAL SULFUR.

- DECISION
  - FOR 200 PPMV SULFUR IN TREATED GAS, USE SELEXOL (BENFIELD HAS FORMATE PROBLEMS).
  - FOR <1 PPMV SULFUR IN TREATED GAS, USE RECTISOL UNLESS A PROVEN COS HYDROLYSIS CATALYST CAN BE FOUND. IN THAT CASE SELEXOL LOOKS SLIGHTLY MORE ATTRACTIVE.
  - SULFINOL HAS LOOKED ATTRACTIVE IN PAST STUDIES AND SHOULD BE INVESTIGATED WHEN MORE DETAILED DESIGN WORK IS DONE.

Table 3.4

Methanation Evaluation & Decision

- EVALUATION
  - HOT RECYCLE FIXED-BED METHANATION USES COMMERCIALLY AVAILABLE CATALYSTS AND EQUIPMENT (EXCEPT POSSIBLY HOT RECYCLE COMPRESSOR) AND RECOVERS MORE HIGH-LEVEL ENERGY THAN OTHER SCHEMES.
  - CAPITAL COST DIFFERENCES BETWEEN HOT AND COLD-RECYCLE METHANATORS IS NOT SIGNIFICANT.
  - DESIGN DONE BY TEAM MEMBERS (FE-2240-31) IS AVAILABLE FOR USE AS A STARTING BASE.

- DECISION
  - USE COLD RECYCLE METHANATION SCHEME.

    (DUE TO SMALL DIFFERENCES BETWEEN SYSTEMS AND TEAM'S FAMILIARITY WITH PREVIOUS DESIGNS)

Table 3.5

Compression & Drying
Evaluation & Decision

- EVALUATION
  - COMPRESSORS MAY BE LOCATED AT SEVERAL PLACES

	ADVANTAGES	DISADVANTAGES
- AHEAD OF AGR	LOWERS AGR COST	MUST COMPRESS SOUR GAS AND CO_2
- BETWEEN AGR AND METHANATION	LOWERS METHANATION COST, LESS GAS TO COMPRESS AFTER CO_2 REMOVAL	COMPRESS H_2+CO NOT CH_4
- AFTER METHANATION	COMPRESS 1/2 VOLUME OF GAS	

  - DRYING IS A LOW-COST SYSTEM AND NEEDS TO BE THE LAST IN THE GAS PURIFICATION TRAIN.
- DECISION
  - FOR METHANE, COMPRESS TO 600 PSIG AHEAD OF AGR, THEN COMPRESS TO 1000 PSIG AFTER METHANATION.

CO Shift: System 20A

Using the assumption of a 50°F approach to equilibrium, the quantity of CO required to be shifted to achieve an H_2/CO ratio of 3.1, in the total gas feeding the methanation system was calculated. From this, the bypass gas quantity was calculated and the material balance defined for the streams in the shift conversion system.

The process simulation program was run to define enthalpy curves for the various process streams in the system. The heat exchange train was then designed to preheat the inlet gas, cool the effluent, and recover the heat of reaction as steam. Based on space velocities from the data base[10] and heat transfer design parameters available in-house, the individual equipment items in the CO shift system were rough-sized to define the necessary parameters for equipment costing.[3] Following the design of the system, an equipment list and process flow diagrams were prepared.

Acid Gas Removal: System 4

The mass balance for the Rectisol system was calculated based on the percentage removals used in the reference design. This allowed the establishment of design parameters which were ratioed to the corresponding elements in the reference design to develop factors for prorating utilities consumption. As discussed earlier, both the reference system design and capital investment were adjusted. This was made possible because the individual equipment items were costed in detail in this design and their utilities consumptions identified as to utilization within the system.

The cost of the facility was estimated by an exponential cost capacity relationship, with the capacity parameter being the total moles of acid gas removed; and the capacity exponent set equal to 0.6 based on previous experience with such units.

Methanation: System 20B

Since the reference design for the methanation system called for an operating pressure of approximately 300 psia, the possibility existed to operate the MBG upgrading plant with an inlet MBG pressure of less than 600 psia. This could have been done by eliminating one of the raw gas compression stages in the K-T System 7, compression, or by operating the Texaco gasifiers at lower pressure. To simplify the design task and to provide for maximum interchangeability and flexibility in the facility design, it was decided to leave the MBG Facility design intact, and to employ a hot gas expander in the methanation system to recover the pressure energy of the inlet gas. The energy thus recovered is used to drive the recycle gas compressors which discharge at a pressure of 1015 psia at the plant battery limits. The use of higher pressure methanation is subject to commercial demonstration of a methanation catalyst at that pressure. The pressure level used in the study appears to be consistent with levels at which commercial methanation catalysts have been demonstrated.

The methanation reaction yields were estimated based on those shown in the reference design and a material balance was then prepared based on these yields. Using the process simulation program, enthalpy curves were developed for the internal process streams, and a heat exchanger network was developed to raise steam, preheat the feed gas, and cool the effluent. A pressure profile was estimated for the system and the process simulator was used to compute horsepower for the expander, recycle gas compressor, and product gas compressor.

A trace sulfur removal system using zinc oxide was included ahead of the methanation reactors and a product gas drying system was specified for the cooled compressor discharge gas. The bases for these two designs are found in Reference 11.

Following the establishment of the equipment configuration and energy/material balance, the equipment was rough-sized using design parameters from the reference design and in-house experience. As with the CO shift system, equipment lists and process flow diagrams were prepared.

Utility System Impacts

After the individual MBG upgrading plant systems were designed, the system utilities requirements were totaled and reduced to net values to determine their impact on the MBG facility. As stated earlier, the primary impacts were in System 15, Steam Generation/Distribution, and System 18, Waste Water Treatment. Capital and operating costs for these systems were recalculated based on these incremental impacts.

4.0 DISCUSSION

This section presents a discussion of the two MBG upgrading plant designs resulting from the study task. The configuration of the designs is such that it is more illustrative to describe first the individual modules associated with K-T and Texaco gasification, and then to show how the combination of modules in each design impacts the remainder of the MBG facility.

4.1 Koppers-Totzek MBG Upgrading Module Process Description

The MBG upgrading module consists of:

- System 7, Compression
- System 20A, CO Shift
- System 4, Acid Gas Removal
- System 20B, Methanation.

Drawing 57-02 is a block flow diagram showing the arrangement of these systems within the module. Table 4.1 presents the module material balance for the Koppers-Totzek SNG module.

System 7: Compression

The configuration of this system, as a result of the tradeoff analyses, was left as it is in the K-T Reference Facility Design. The system consists of two stages of axial compression followed by two stages of centifugal compression. For one module, approximately 11.89 MMSCFH of cooled raw gas from System 3, Gas Cleanup and Cooling, is compressed from 14.9 psia to about 700 psia. The total compressor horsepower required is about 84,400 HP.

System 20A: CO Shift

Drawing 57-06 is a process flow diagram of this system. Table 4.2 contains the associated material balance. 29,083 moles per hour of raw gas from System 7, Compression enters the CO shift system at 690 psia, 100°F. Approximately 25 percent of the inlet gas is bypassed around the shift conversion step, passing instead through COS hydrolysis. The quantity of bypass gas is controlled by an H_2/CO ratio analyzer

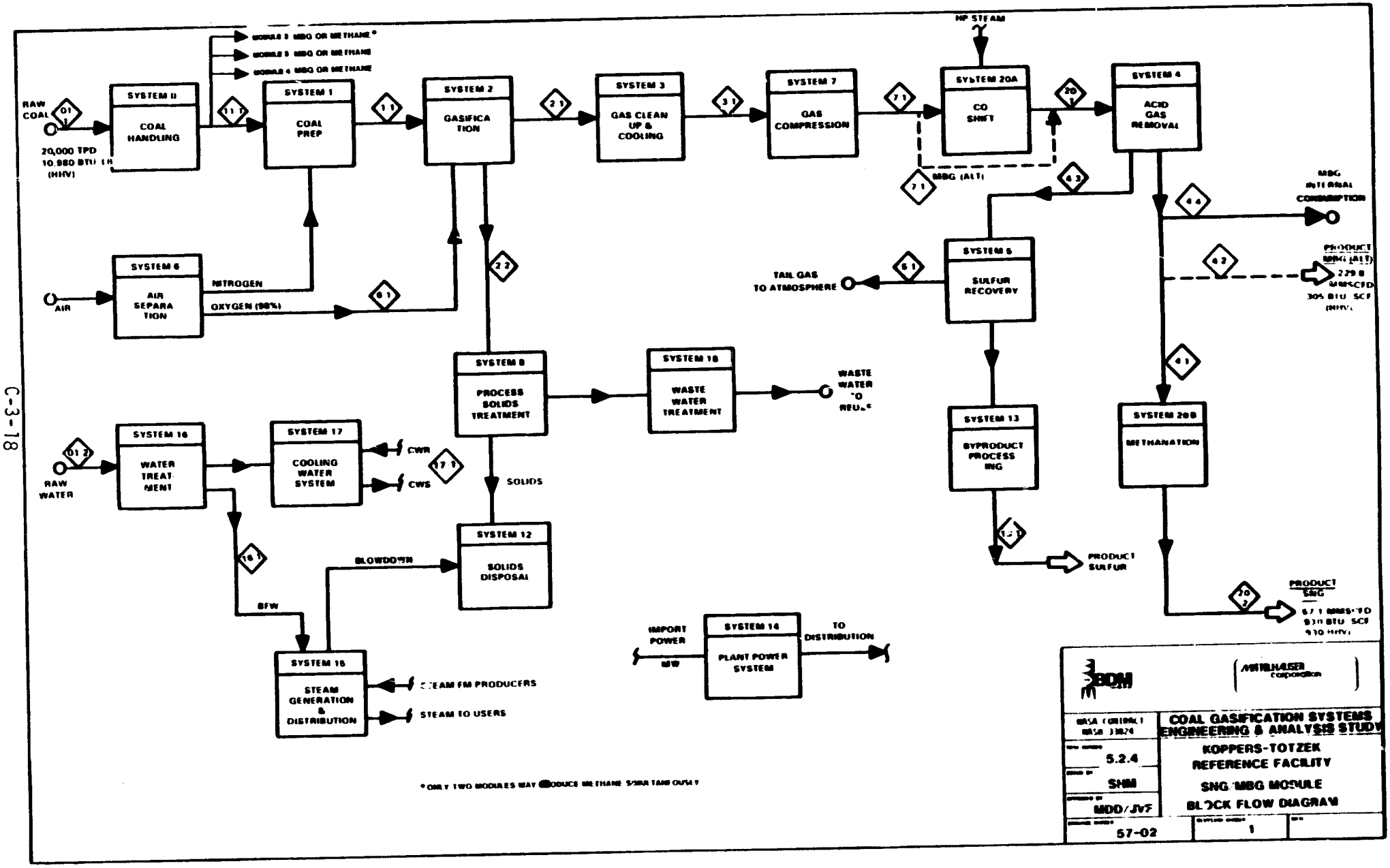

MODULE 3 MBG OR METHANE *
MODULE 5 MBG OR METHANE
MODULE 4 MBG OR METHANE
HP STEAM
RAW COAL
20,000 TPD
10,980 BTU LB
(HHV)
SYSTEM 11
COAL HANDLING
SYSTEM 1
COAL PREP
SYSTEM 2
GASIFICATION
SYSTEM 3
GAS CLEAN UP & COOLING
SYSTEM 7
GAS COMPRESSION
SYSTEM 20A
CO SHIFT
SYSTEM 4
ACID GAS REMOVAL
MBG (ALT)
MBG INTERNAL CONSUMPTION
PRODUCT MBG (ALT)
229 B MMSCFD
305 BTU SCF (HHV)
AIR
SYSTEM 6
AIR SEPARATION
NITROGEN
OXYGEN (98%)
TAIL GAS TO ATMOSPHERE
SYSTEM 5
SULFUR RECOVERY
SYSTEM 8
PROCESS SOLIDS TREATMENT
SYSTEM 18
WASTE WATER TREATMENT
WASTE WATER TO REUSE
SYSTEM 13
BYPRODUCT PROCESSING
SYSTEM 20B
METHANATION
PRODUCT SULFUR
PRODUCT SNG
RAW WATER
SYSTEM 16
WATER TREATMENT
SYSTEM 17
COOLING WATER SYSTEM
CWR
CWS
SOLIDS
SYSTEM 12
SOLIDS DISPOSAL
BLOWDOWN
BFW
SYSTEM 15
STEAM GENERATION & DISTRIBUTION
STEAM FM PRODUCERS
STEAM TO USERS
IMPORT POWER
MW
SYSTEM 14
PLANT POWER SYSTEM
TO DISTRIBUTION
* ONLY TWO MODULES MAY PRODUCE METHANE SIMULTANEOUSLY
BOM
MITTELHAUSER CORPORATION
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
SHM
MDD/JVF
KOPPERS-TOTZEK REFERENCE FACILITY
SNG/MBG MODULE
BLOCK FLOW DIAGRAM
57-02
1

Table 4.1

Material Balance
Koppers-Totzek Process
Integrated SNG/Module Definition

STREAM NUMBER			01-1		01-2		1-1	
STREAM ID			RAW COAL FEED		RAW WATER		COAL TO GASIFIER	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01		253,637				253,637
HYDROGEN	H_2	2.016		17,925				17,925
OXYGEN	O_2	32.0		23,901				23,901
NITROGEN & ARGON	N_2	28.016		5,761				5,761
SULFUR	S	32.06		15,449				15,449
CHLORINE	Cl	35.453		494				494
CARBON MONOXIDE	CO	28.01		-				-
CARBON DIOXIDE	CO_2	44.01		-				-
METHANE	CH_4	16.042		-				-
ETHYLENE	C_2H_4	28.052		-				-
ETHANE	C_2H_6	30.068		-				-
PROPYLENE	C_3H_6	42.078		-				-
PROPANE	C_3H_8	44.094		-				-
HYDROGEN SULFIDE	H_2S	34.076		-				-
CARBONYL SULFIDE	COS	60.075		-				-
CARBON DISULFIDE	CS_2	76.13		-				-
SULFUR DIOXIDE	SO_2	64.06		-				-
NITROUS OXIDE	NO	30.008		-				-
AMMONIA	NH_3	17.031		-				-
HYDROGEN CYANIDE	HCN	27.026		-				-
HYDROGEN CHLORIDE	HCl	36.461		-				-
NAPHTHA	-			-				59,650
TAR & OIL	-			59,650				-
ASH	-			-				376,617
OTHER SOLIDS				376,817				3,606
SUBTOTAL, DRY				39,853				880,623
WATER	H_2O	18.016		416,670				-
TOTAL, WET			-	-				-
GAS MOLECULAR WEIGHT			-					140
TEMPERATURE, °F			-					-
PRESSURE, PSIA			-					-
SCFH					2,200			
GPM								

Table 4.1 (Continued)

Material Balance
Koppers-Totzek Process
Integrated SNG/Module Definition

STREAM NUMBER			2-1		2-2		3-1	
STREAM ID			RAW GAS FROM GASIFIER		ASH FM GASIFIER		COOL GAS FROM GASIFIER	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01	-	-		-	-	-
HYDROGEN	H_2	2.016	8,125.4	16,380		-	8,125.4	16,381
OXYGEN	O_2	32.0	-	-		-	-	-
NITROGEN & ARGON	N_2	28.016	419.9	11,764		-	419.9	11,764
SULFUR	S	32.06	-	-		-	-	-
CHLORINE	Cl	35.453	-	-		-	-	-
CARBON MONOXIDE	CO	28.01	17,405.8	487,536		-	17,405.8	487,536
CARBON DIOXIDE	CO_2	44.01	2,483.2	109,286		-	2,483.2	109,286
METHANE	CH_4	16.042	125	2,005		-	125	2,005
ETHYLENE	C_2H_4	28.052	-	-		-	-	-
ETHANE	C_2H_6	30.068	-	-		-	-	-
PROPYLENE	C_3H_6	42.078	-	-		-	-	-
PROPANE	C_3H_8	44.094	-	-		-	-	-
HYDROGEN SULFIDE	H_2S	34.076	433.2	14,762		-	433.2	14,762
CARBONYL SULFIDE	COS	60.075	48.7	2,926		-	48.7	2,926
CARBON DISULFIDE	CS_2	76.13	-	-		-	-	-
SULFUR DIOXIDE	SO_2	64.06	-	-		-	-	-
NITROUS OXIDE	NO	30.008	-	-		-	-	-
AMMONIA	NH_3	17.031	-	-		-	-	-
HYDROGEN CYANIDE	HCN	27.026	-	-		-	-	-
HYDROGEN CHLORIDE	HCl	36.461	14.3	521		-	-	-
NAPHTHA	-		-	-		-	-	-
TAR & OIL	-		-	-		-	-	-
ASH	-		-	-		59,650	-	-
OTHER SOLIDS			-	-		12,684	-	-
SUBTOTAL, DRY			29,055.5	645,180		72,334	29,041.2	644,660
WATER	H_2O	18.016	10,326.5	186,042		217,000	2,244.4	40,435
TOTAL, WET			39,382	831,222		289,336	31,285.6	685,095
GAS MOLECULAR WEIGHT			21.1		-		21.9	
TEMPERATURE, °F			350		-		100	
PRESSURE, PSIA			15.4		-		14.9	
SCFH			14,952,083		-		11,878,140	
GPM			-		433		-	

Table 4.1 (Continued)

Material Balance
Koppers-Totzek Process
Integrated SNG/Module Definition

STREAM NUMBER			4-1		4-4		4-3	
STREAM ID			SWEET GAS TO METHANATION (100% SNG PRODUCTION)		PRODUCT NBG (100% NBG PRODUCTION)		ACID GAS TO SULFUR RECOVERY	
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR
CARBON	C	12.01	-	-	-	-	-	-
HYDROGEN	H_2	2.016	17,134.2	34,542.5	7,490.4	15,101	-	-
OXYGEN	O_2	32.0	-	-	-	-	-	-
NITROGEN & ARGON	N_2	28.016	360.2	10,089	387.3	10,848	-	-
SULFUR	S	32.06	-	-	-	-	-	-
CHLORINE	Cl	35.453	-	-	-	-	-	-
CARBON MONOXIDE	CO	28.01	5,509.9	154,332	16,023.3	448,813	-	-
CARBON DIOXIDE	CO_2	44.01	231.1	10,171	1,200.6	52,838	2,676.3	117,784
METHANE	CH_4	16.042	107.2	1,719	115.3	1,849	-	-
ETHYLENE	C_2H_4	28.052	-	-	-	-	-	-
ETHANE	C_2H_6	30.068	-	-	-	-	-	-
PROPYLENE	C_3H_6	42.078	-	-	-	-	-	-
PROPANE	C_3H_8	44.094	-	-	-	-	-	-
HYDROGEN SULFIDE	H_2S	34.076	TR	TR	0.977	33.3	479.1	16,328
CARBONYL SULFIDE	COS	60.075	TR	TR	4.08	245.1	2.3	138
CARBON DISULFIDE	CS_2	76.13	-	-	-	-	-	-
SULFUR DIOXIDE	SO_2	64.06	-	-	-	-	-	-
NITROUS OXIDE	NO	30.008	-	-	-	-	-	-
AMMONIA	NH_3	17.031	-	-	-	-	-	-
HYDROGEN CYANIDE	HCN	27.026	-	-	-	-	-	-
HYDROGEN CHLORIDE	HCl	36.461	-	-	-	-	-	-
NAPHTHA	-		-	-	-	-	-	-
TAR & OIL	-		-	-	-	-	-	-
ASH	-		-	-	-	-	-	-
OTHER SOLIDS			-	-	-	-	-	-
SUBTOTAL, DRY			23,342.6	210,854	25,221.9	529,727	3,157.7	134,250
WATER	H_2O	18.016	1.3	23	3.7	67	89.8	1,618
TOTAL, WET			23,343.9	210,877	25,225.6	529,794	3,247.5	135,868
GAS MOLECULAR WEIGHT			9.03		21.0		41.8	
TEMPERATURE, °F			100		60		120	
PRESSURE, PSIA			297		615		20	
SCFH			8,862,930		9,574,629		1,232,971	
GPM			-		-		-	

Table 4.1 (Continued)

Material Balance
Koppers-Totzek Process
Integrated SNG/Module Definition

STREAM NUMBER			4-2		4-2		5-1	
STREAM ID			MBG INTERNAL USAGE(100 MBG CASE)		~~MBG INTERNAL CONSUMPTION (100% SNG CASE)~~		TAIL GAS FROM SULFUR RECOVERY	
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR
CARBON	C	12.01	-	-	-	-	-	-
HYDROGEN	H_2	2.016	631.7	1,273	2,014.7	4,062	-	-
OXYGEN	O_2	32.0	-	-	-	-	24.6	49
NITROGEN & ARGON	N_2	28.016	32.6	913	42.3	1,185	954	26,727
SULFUR	S	32.06	-	-	-	-	-	-
CHLORINE	Cl	35.453	-	-	-	-	-	-
CARBON MONOXIDE	CO	28.01	1,351.2	37,847	647.9	18,148	-	-
CARBON DIOXIDE	CO_2	44.01	103.7	4,564	27.2	1,197	1,175.3	51,725
METHANE	CH_4	16.042	9.7	155	12.6	202	-	-
ETHYLENE	C_2H_4	28.052	-	-	-	-	-	-
ETHANE	C_2H_6	30.068	-	-	-	-	-	-
PROPYLENE	C_3H_6	42.078	-	-	-	-	-	-
PROPANE	C_3H_8	44.094	-	-	-	-	-	-
HYDROGEN SULFIDE	H_2S	34.076	0.08	2.7	TR	-	TR	TR
CARBONYL SULFIDE	COS	60.075	0.34	20.4	TR	-	TR	TR
CARBON DISULFIDE	CS_2	76.13	-	-	-	-	-	-
SULFUR DIOXIDE	SO_2	64.06	-	-	-	-	-	-
NITROUS OXIDE	NO^2	30.008	-	-	-	-	-	-
AMMONIA	NH_3	17.031	-	-	-	-	-	-
HYDROGEN CYANIDE	HCN	27.026	-	-	-	-	-	-
HYDROGEN CHLORIDE	HCl	36.461	-	-	-	-	-	-
NAPHTHA	-		-	-	-	-	-	-
TAR & OIL	-		-	-	-	-	-	-
ASH	-		-	-	-	-	-	-
OTHER SOLIDS			-	-	-	-	-	-
SUBTOTAL, DRY			2,129.3	44,776	2,744.7	24,793	2,153.9	78,502
WATER	H_2O	18.016	0.77	13.9	0.16	2.9	147.3	2,654
TOTAL, WET			2,130.1	44,790	2,744.9	24,796	2,301.2	81,156
GAS MOLECULAR WEIGHT			21.0		9.03		35.3	
TEMPERATURE, °F			60		100		100	
PRESSURE, PSIA			315		297		14.4	
SCFH			808,731		1,042,150		873,443	
GPM			-		-		-	

Table 4.1 (Continued)

Material Balance
Koppers-Totzek Process
Integrated SNG/Module Definition

STREAM NUMBER			6-1		7-1		15-1	
STREAM ID			OXIDANT FEED TO GASIFIER		COMPRESSED RAW GAS TO SHIFT		SULFUR BY-PRODUCT	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01	-	-	-	-	-	-
HYDROGEN	H_2	2.016	-	-	8,125.4	16,381	-	-
OXYGEN	O_2	32.0	10,501.4	336,045	-	-	-	-
NITROGEN & ARGON	N_2	28.016	214.3	6,004	419.9	11,764	-	-
SULFUR	S	32.06	-	-	-	-	481.9	15,450
CHLORINE	Cl	35.453	-	-	-	-	-	-
CARBON MONOXIDE	CO	28.01	-	-	17,405.8	487,536	-	-
CARBON DIOXIDE	CO_2	44.01	-	-	2,483.2	109,286	-	-
METHANE	CH_4	16.042	-	-	125	2,005	-	-
ETHYLENE	C_2H_4	28.052	-	-	-	-	-	-
ETHANE	C_2H_6	30.068	-	-	-	-	-	-
PROPYLENE	C_3H_6	42.078	-	-	-	-	-	-
PROPANE	C_3H_8	44.094	-	-	-	-	-	-
HYDROGEN SULFIDE	H_2S	34.076	-	-	433.2	14,762	-	-
CARBONYL SULFIDE	COS	60.075	-	-	48.7	2,926	-	-
CARBON DISULFIDE	CS_2	76.13	-	-	-	-	-	-
SULFUR DIOXIDE	SO_2	64.06	-	-	-	-	-	-
NITROUS OXIDE	NO	30.008	-	-	-	-	-	-
AMMONIA	NH_3	17.031	-	-	-	-	-	-
HYDROGEN CYANIDE	HCN	27.026	-	-	-	-	-	-
HYDROGEN CHLORIDE	HCl	36.461	-	-	-	-	-	-
NAPHTHA	-		-	-	-	-	-	-
TAR & OIL	-		-	-	-	-	-	-
ASH	-		-	-	-	-	-	-
OTHER SOLIDS			-	-	-	-	-	-
SUBTOTAL, DRY			10,715.7	342,049	29,041.2	644,660	481.9	15,450
WATER	H_2O	18.016	-	-	41.6	749	-	-
TOTAL, WET			10,715.7	342,049	29,082.8	645,409	481.9	15,450
GAS MOLECULAR WEIGHT			31.92		22.2		-	
TEMPERATURE, °F			140		100		60	
PRESSURE, PSIA			21.6		700		-	
SCFH			4,068,408		11,041,832		-	
GPM			-		-		-	

Table 4.1 (Concluded)

Material Balance
Koppers-Totzek Process
Integrated SNG/Module Definition

STREAM NUMBER			16-1		20-1		20-2	
STREAM ID			MAKE-UP WATER TO BFW SYSTEM		SHIFTED GAS TO ACID GAS REMOVAL		PRODUCT SNG(100 SNG PRODUCTION)	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01			-	-	-	-
HYDROGEN	H_2	2.016			19,279.3	38,867	58.1	117
OXYGEN	O_2	32.0			-	-	-	-
NITROGEN & ARGON	N_2	28.016			419.9	11,764	360.1	10,086
SULFUR	S	32.06			-	-	-	-
CHLORINE	Cl	35.453			-	-	-	-
CARBON MONOXIDE	CO	28.01			6,233.8	174,609	1.6	45
CARBON DIOXIDE	CO_2	44.01			13,701.1	602,985	93.3	4,106
METHANE	CH_4	16.042			125	2,005	5,753.3	92,283
ETHYLENE	C_2H_4	28.052			-	-	-	-
ETHANE	C_2H_6	30.068			-	-	-	-
PROPYLENE	C_3H_6	42.078			-	-	-	-
PROPANE	C_3H_8	44.094			-	-	-	-
HYDROGEN SULFIDE	H_2S	34.076			479.2	16,331	TR	TR
CARBONYL SULFIDE	COS	60.075				163	TR	TR
CARBON DISULFIDE	CS_2	76.13			-	-	-	-
SULFUR DIOXIDE	SO_2	64.06			-	-	-	-
NITROUS OXIDE	NO	30.008			-	-	-	-
AMMONIA	NH_3	17.031			-	-	-	-
HYDROGEN CYANIDE	HCN	27.026			-	-	-	-
HYDROGEN CHLORIDE	HCl	36.461			-	-	-	-
NAPHTHA	-				-	-	-	-
TAR & OIL	-				-	-	-	-
ASH	-				-	-	-	-
OTHER SOLIDS					-	-	-	-
SUBTOTAL, DRY					40,241.0	846,724	6,266.4	106,637
WATER	H_2O	18.016			71.6	1,290	0.93	16
TOTAL, WET					40,312.6	848,014	6,267.3	106,653
GAS MOLECULAR WEIGHT					21.0		17.0	
TEMPERATURE, °F					100		100	
PRESSURE, PSIA					700		1,030	
SCFH					15,305,402		2,379,492	
GPM			1,404		-		-	

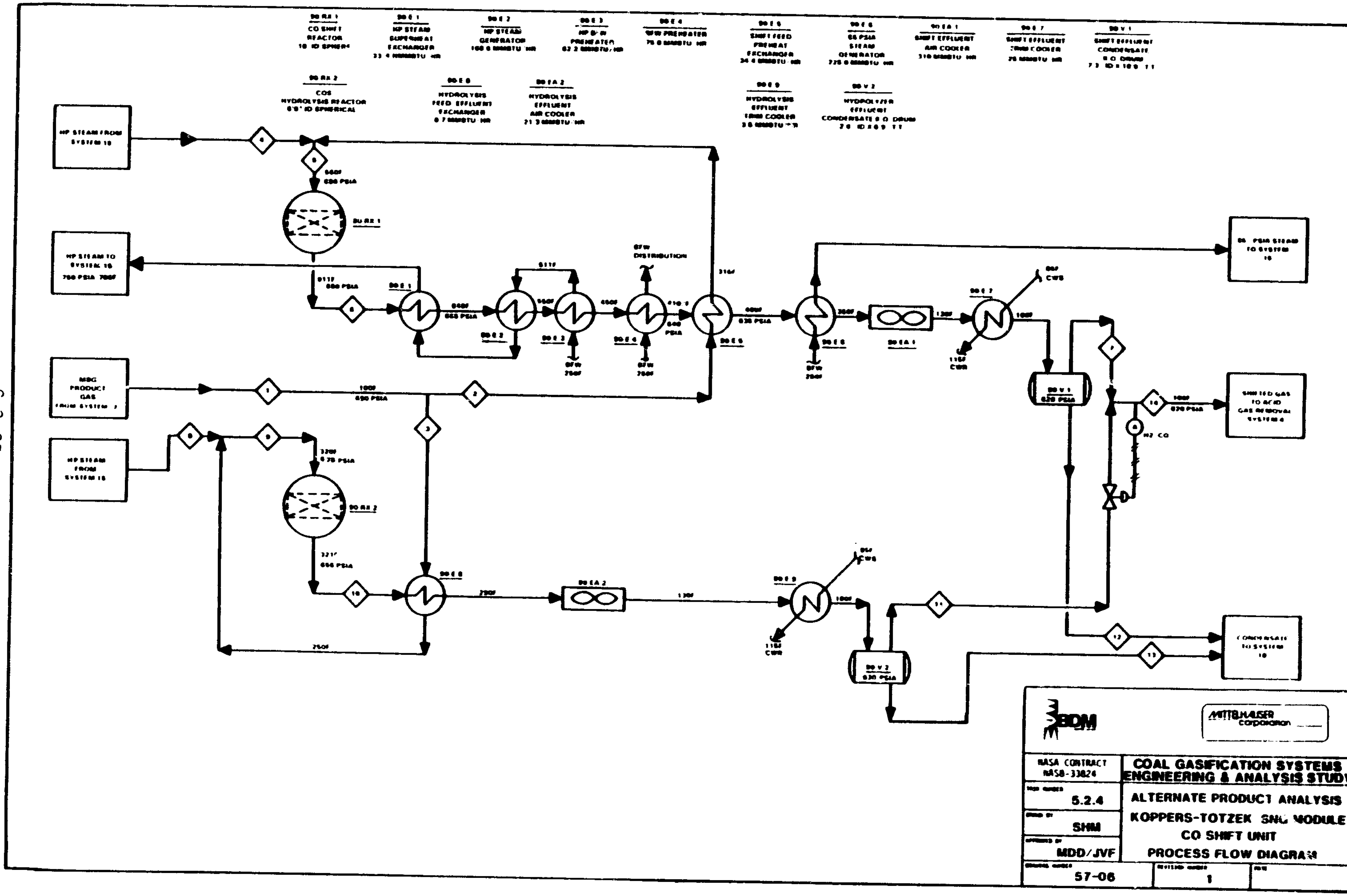

HP STEAM FROM SYSTEM 10
HP STEAM TO SYSTEM 10 700 PSIA 700F
MBG PRODUCT GAS FROM SYSTEM 7
HP STEAM FROM SYSTEM 10
65 PSIA STEAM TO SYSTEM 10
SHIFTED GAS TO ACID GAS REMOVAL SYSTEM 8
CONDENSATE TO SYSTEM 10
90 RX 1 CO SHIFT REACTOR
90 RX 2 COS HYDROLYSIS REACTOR
90 E 1 HP STEAM SUPERHEAT EXCHANGER
90 E 2 HP STEAM GENERATOR
90 E 3 HP BFW PREHEATER
90 E 4 BFW PREHEATER
90 E 5 SHIFT FEED PREHEAT EXCHANGER
90 E 6 65 PSIA STEAM GENERATOR
90 EA 1 SHIFT EFFLUENT AIR COOLER
90 E 7 SHIFT EFFLUENT TRIM COOLER
90 V 1 SHIFT EFFLUENT CONDENSATE K.O. DRUM
90 E 8 HYDROLYSIS FEED EFFLUENT EXCHANGER
90 EA 2 HYDROLYSIS EFFLUENT AIR COOLER
90 E 9 HYDROLYSIS EFFLUENT TRIM COOLER
90 V 2 HYDROLYSIS EFFLUENT CONDENSATE K.O. DRUM
BFW DISTRIBUTION
H2 CO
BDM
MITTELHAUSER corporation
NASA CONTRACT NAS8-33824
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
ALTERNATE PRODUCT ANALYSIS
KOPPERS-TOTZEK SNG MODULE
CO SHIFT UNIT
SHM
PROCESS FLOW DIAGRAM
MDD/JVF
57-06
1

Table 4.2

Material Balance
Koppers-Totzek SNG Module
CO Shift Unit

STREAM NUMBER			1		2		3	
STREAM ID			MBG FROM COMPRESSION		CO SHIFT REACTOR FEED GAS		COS HYDROLYSIS REACTOR FEED GAS	
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	8,125.4		6,094.0		2,031.4	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.216	419.9		314.9		105.0	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	17,405.8		13,054.2		4,351.5	
CARBON DIOXIDE	CO_2	44.01	2,483.2		1,862.4		620.8	
METHANE	CH_4	16.042	125.0		93.7		31.3	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076	433.2		324.9		108.3	
CARBONYL SULFIDE	COS	60.075	48.7		36.5		12.2	
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO_2	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			29,041.2		21,780.6		7,229.8	
WATER	H_2O	18.016	41.6		31.2		10.4	
TOTAL, WET			29,082.8	645,409	21,811.8	484,057	7,240.2	161,352
GAS MOLECULAR WEIGHT			22.29		22.28		22.29	
TEMPERATURE, °F			100		100		100	
PRESSURE, PSIA			690		690		690	
SCFH X 10^6			11.4		8.28		2.76	
GPM								

Table 4.2 (Continued)

Material Balance
Koppers-Totzek SNG Module
CO Shift Unit

STREAM NUMBER			4		5		6	
STREAM ID			HP STEAM TO CO SHIFT REACTOR		CO SHIFT REACTOR INLET		CO SHIFT REACTOR OUTLET	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016			6,094.0		17,265.9	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	8,125.4		314.9		314.9	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01			13,054.2		1,882.3	
CARBON DIOXIDE	CO_2	44.01			1,867.4		13,068.2	
METHANE	CH_4	16.042			93.7		93.7	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076			324.9		358.8	
CARBONYL SULFIDE	COS	60.075			36.5		2.6	
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO_2	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY					21,780.6		33,343.7	
WATER	H_2O	18.016	38,375.9		38,767.1		27,204.0	
TOTAL, WET			38,375.9	697,865	60,547.7	1,181,922	60,547.7	1,181,922
GAS MOLECULAR WEIGHT			18		19.52		19.52	
TEMPERATURE, °F			700		560		911	
PRESSURE, PSIA			750		680		660	
SCFH X 10^6					22.98		22.98	
GPM								

Table 4.2 (Continued)

Material Balance
Koppers-Totzek SNG Module
CO Shift Unit

STREAM NUMBER			7		8		9	
STREAM ID			CO SHIFT REACTOR EFFLUENT TO AGR		HP STEAM TO COS HYDROLYSIS REACTOR		COS HYDROLYSIS REACTOR INLET	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	17,265.9				2,031.4	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	314.9				105.0	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	1,882.3				4,351.5	
CARBON DIOXIDE	CO_2	44.01	13,068.2				620.8	
METHANE	CH_4	16.042	93.7				31.3	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076	358.8				108.3	
CARBONYL SULFIDE	COS	60.075	2.6				12.2	
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO^2	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			33,343.7				7,229.8	
WATER	H_2O	18.016	60.6		1,255.0		1,265.4	
TOTAL, WET			33,404.3	692,907	1,255.0	22,610	8,495.2	183,692
GAS MOLECULAR WEIGHT			20.74		18		21.65	
TEMPERATURE, °F			100		700		320	
PRESSURE, PSIA			620		750		675	
SCFH X 10^6			12.67				3.22	
GPM								

Table 4.2 (Continued)

Material Balance
Koppers-Totzek SNG Module
CO Shift Unit

STREAM NUMBER			10		11		12	
STREAM ID			COS HYDROLYSIS REACTOR OUTLET		COS HYDROLYSIS REACTOR EFFLUENT		SHIFT CONDENSATE TO TREATING	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS TO AGR/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	2,031.4		2,031.4			
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	105.0		105.0			
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	4,351.5		4,351.5			
CARBON DIOXIDE	CO_2	44.01	632.9		632.9			
METHANE	CH_4	16.042	31.3		31.3			
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076	120.4		120.4			
CARBONYL SULFIDE	COS	60.075	0.1		0.1			
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			7,272.6		7,272.6			
WATER	H_2O	18.016	1,222.6		11.0		27,143.4	
TOTAL, WET			8,495.2	183,692	7,283.6	161,864	27,143.4	489,015
GAS MOLECULAR WEIGHT			21.65				18	
TEMPERATURE, °F			321				100	
PRESSURE, PSIA			655				620	
SCFH X 10^6			3.22					
GPM							978	

Table 4.2 (Concluded)

Material Balance
Koppers-Totzek SNG Module
CO Shift Unit

STREAM NUMBER			13		14			
STREAM ID			HYDROLYSIS CONDENSATE TO TREATING		SHIFTED GAS TO ACID GAS REMOVAL			
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016			19,297.3			
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016			419.9			
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01			6,233.8			
CARBON DIOXIDE	CO_2	44.01			13,701.1			
METHANE	CH_4	16.042			125.0			
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076			479.2			
CARBONYL SULFIDE	COS	60.075			2.7			
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY					40,241.0			
WATER	H_2O	18.016	1,211.6		71.6			
TOTAL, WET			1,211.6	21,828	40,312.6	848,014		
GAS MOLECULAR WEIGHT			18		21.0			
TEMPERATURE, °F			100		100			
PRESSURE, PSIA			630		620			
SCFH X 10^6					15.30			
GPM			44					

installed on the combined outlet gas stream going to System 4, Acid Gas Removal. This analyzer operates a bypass flow control valve which maintains the H_2/CO ratio in the effluent gas at 3.1.

The gas which is to be shifted passes first through 90-E-5, in which it is heated to 310°F against shift converter effluent gas. The preheated gas is then mixed with 735 psig, 700°F superheated steam to raise the steam-to-dry gas ratio to 1.08. The temperature of the mixed gas stream is also raised to 560°F as required for inlet to the shift converter.

The gas next passes through 90-V-1, CO shift reactor. This is a spherical reactor containing a fixed bed of cobalt-molybdate catalyst, of approximately 24,400 cubic feet in volume. Here, the CO shift reaction takes place.

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

A 50°F temperature approach to equilibrium has been assumed. This results in a CO conversion of 85.5% based on reactor feed. For the purposes of this design it has also been assumed that the COS hydrolysis reaction occurs over the catalyst.

$$COS + H_2O \rightleftharpoons CO_2 + H_2S$$

A 50°F approach to equilibrium has been assumed. At the outlet temperature of the shift converter, this corresponds to a 93.7 percent conversion of COS based on shift reactor feed.

The exothermic shift and hydrolysis reactions cause the shifted gas to leave the reactor at about 911°F. The gas is cooled in a series of shell-and-tube heat exchangers summarized in Table 4.3. In these exchangers, the shift reaction heat is recovered as useful energy in the form of superheated steam which is exported to System 15, Steam Generation and Distribution.

Table 4.3

CO Shift Effluent Cooling Train

ITEM NUMBER	NAME	DUTY 10^6 BTU/HR	EFF. ΔT °F	U	AREA FT^2
E-1	HP STEAM SUPERHEAT EXCHANGER	33.4	265.6	60	2096
E-2	HP STEAM GENERATOR #1	166.6	136.0	120	10208
E-3	HP BFW PREHEATER #1	62.2	98.5	80	7893
E-4	HP BFW PREHEATER #2	75.0	94.6	80	9910
E-5	SHIFT FEED PREHEAT EXCHANGER	34.4	182.0	100	1890
E-6	65 PSIA STEAM GENERATOR	225 X 10^6	101.0	120	18564
EA-1	SHIFT EFFLUENT AIR COOLER	310 X 10^6	126.5	100	24506
E-7	SHIFT EFFLUENT TRIM COOLER	25 x 10^6	10		
E-8	HYDROLYSIS FEED/EFFLUENT EXCHANGER	8.7	120.9	60	1199
EA-2	HYDROLYSIS EFFLUENT AIR COOLER	21.3	105.6	100	2017
E-9	HYDROLYSIS EFFLUENT TRIM COOLER	3.5	10	80	4375

The shift bypass gas is preheated to 250°F in 90-E-8, Hydrolysis Feed/Effluent Exchanger, against the hydrolysis reactor effluent. Then, 735 psig, 700°F steam is injected to bring the steam-to-dry gas ratio in the hydrolysis reactor feed to approximately 0.2 and the reactor inlet temperature to 320°F the same as in the reference design.[10] At this temperature, the conversion of COS is calculated to be about 99.2% based on reactor feed. Because of the small amount of COS being hydrolyzed, the temperature rise is only about 1 degree F.

90-V-2, the COS hydrolyzer, is a spherical reactor filled with 575 cubic feet of catalyst. It is assumed for the purposes of this study that a catalyst for this source is commercially available; although this has not yet been demonstrated. The Muscle Shoals demonstration plant contains a COS hydrolysis catalyst which may be suitably demonstrated in the near future. Although the operating conditions for this catalyst are somewhat different from those used in this study[12], the overall plant design and economics are not expected to be materially affected by this difference.

After leaving 90-V-2, the reactor effluent is cooled by preheating feed gas in 90-E-8, and is further cooled to 100°F in 90-EA-2, Hydrolysis Effluent Air Cooler, and 90-E-9, Hydrolysis Effluent Trim Cooler. The gas is then mixed with the cooled shift effluent gas and flows to System 4, Acid Gas Removal.

System 4: Acid Gas Removal

The purpose of this system is to:

- Remove sulfur compounds from the process gas to a total concentration of 1 ppmv or below.
- Remove CO_2 from the process gas to a concentration of approximately 0.5 mole percent.

The process used is the Rectisol process, licensed by Lurgi. The design used for this study is based on a published design.[13]

The Rectisol process uses methanol to accomplish the desired removal of acid gases. As designed for this application, the following seven subsystems would be included in the design:

- Gas Cooling
- Methanol Recover
- Gas Purification
- Flash Regeneration
- H_2S Concentration
- Hot Regeneration
- Refrigeration

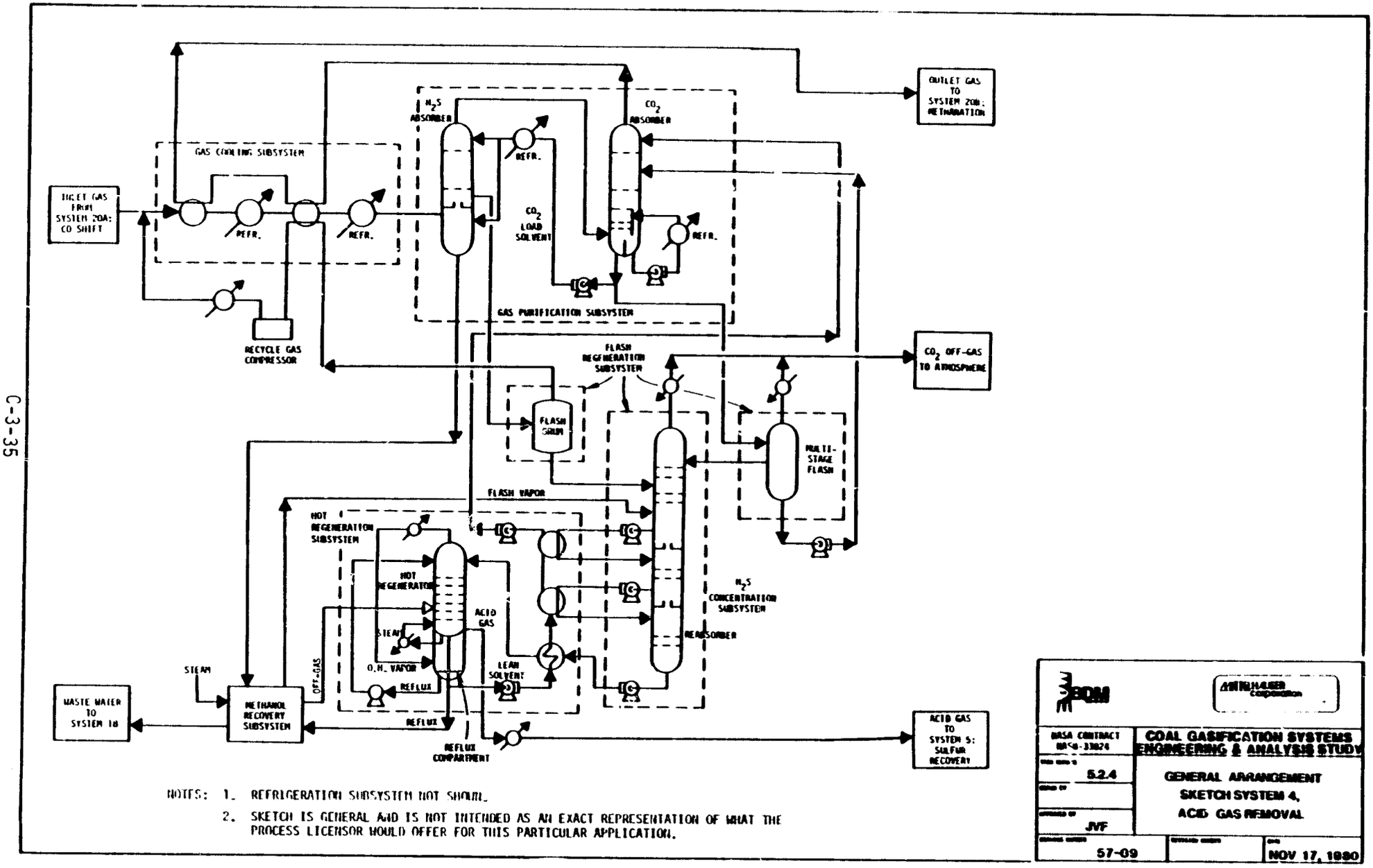
Drawing 57-09 is a representation of the Rectisol System.

The internal details of stream flows, temperatures, and pressures are proprietary and were not calculated for this design. As mentioned earlier, this design was prorated from the reference design on a system level.

The purpose of the gas cooling subsystem is to cool the raw inlet gas from 100°F to subzero temperatures. This is accomplished in a series of shell-and-tube heat exchangers. Part of the water present in the inlet gas condenses in these exchangers and is sent to the methanol recovery susbystem. The cooled gas is sent to the gas purification subsystem.

The methanol recovery subsystem is used to separate the water condensed from the inlet gas from the methanol solvent. There is some question as to whether this system is needed if no condensable light hydrocarbons are present in the inlet gas. However, to provide some conservatism in the cost estimate, it has been retained.

The gas purification subsystem is where the H_2S, organic sulfur compounds, and most of the CO_2 are removed from the gas before the gas is sent to System 20B, Methanation. The cooled gas from the gas cooling subsystem first enters the prewash section of the H_2S Absorber, where it is washed with a small stream of cold methanol to remove organic sulfur compounds. The liquid from the prewash section, containing water



NOTES: 1. REFRIGERATION SUBSYSTEM NOT SHOWN.

2. SKETCH IS GENERAL AND IS NOT INTENDED AS AN EXACT REPRESENTATION OF WHAT THE PROCESS LICENSOR WOULD OFFER FOR THIS PARTICULAR APPLICATION.

BOM	Ambauer Corporation
NASA CONTRACT NAS8-33824	COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4	GENERAL ARRANGEMENT SKETCH SYSTEM 4, ACID GAS REMOVAL
JVF	
57-09	NOV 17, 1980

condensed from the gas in the gas cooling subsystem, flows to the methanol recovery subsystem. The prewashed gas enters the top section of the H_2S Absorber where the H_2S is absorbed. The methanol for the absorption step comes from the CO_2 Absorber, pumped through a refrigerant-cooled heat exchanger. The rich methanol from the base of the upper section of the H_2S Absorber flows to the flash regeneration subsystem. The overhead gas from the H_2S Absorber flows to the CO_2 Absorber.

Treatment in the trayed CO_2 Absorber is done in 3 steps. In the bottom section of the tower, the heat of absorption of the CO_2 is removed by a pump around cooler. In the middle section, the bulk of the CO_2 is removed by countercurrent contact with flash regenerated methanol from the flash regeneration subsystem. In the top section, the final amounts of sulfur compounds and CO_2 are removed by countercurrent contact with lean methanol supplied from the hot regeneration subsystem. Rich methanol, in excess of that needed to supply the H_2S Absorber, flows to the flash regeneration subsystem. The treated gas leaving the top of the CO_2 Absorber is reheated by cooling the incoming gas in the gas cooling subsystem, after which it flows to System 20B, Methanation.

In the flash regeneration subsystem, light gases are recovered from the H_2S rich methanol, and CO_2 rich methanol from the CO_2 Absorber is partly regenerated for use in the middle section of the CO_2 Absorber. Flash gas from the H_2S rich methanol is recompressed and recycled to the inlet gas stream. The CO_2 rich methanol is flashed in several steps. Vacuum blowers compress vacuum flash vapors which are then vented to the atmosphere. The flashed methanol not recycled to the CO_2 Absorber is fed to the H_2S concentration subsystem.

The H_2S concentration subsystem separates CO_2 from the H_2S rich methanol stream, so that the H_2S concentration in the acid gas feeding the sulfur recovery system will be as high as possible, hopefully between 25 and 50 volume percent.

In this subsystem, flashed H_2S rich methanol enters the lower part of the top section of the Reabsorber where it is flashed. The liquid from this "cold flash" is pumped through a shell and tube exchanger where it is heated against hot regenerated methanol and returned to the tower to be flashed again. The liquid from this "warm flash" is pumped through a second exchanger where it is heated a second time against hot regenerated methanol and returned to the tower to be flashed a third time.

The gases from the three flashes rise in the tower and are washed in the top section of the tower by flashed CO_2 rich methanol from the flash regeneration subsystem. The washing reabsorbs any flashed hydrogen sulfide and part of the carbonyl sulfide. In the liquid leaving the bottom of the Reabsorber the dissolved gas is up to 50 volume percent H_2S. This liquid is pumped to the hot regeneration subsystem where the acid (Claus) gas is stripped from the methanol.

An extra connection from the acid gas line to below tray one in the Reabsorber is provided to permit production of acid gas of about 50 volume percent H_2S even when the sulfur content of the mixed gas is lower than expected. Overhead gas from the Reabsorber, rich in CO_2, is vented to the atmosphere.

The purpose of the hot regeneration subsystem is to strip the carbon dioxide, non-methane hydrocarbons, hydrogen sulfide and carbonyl suflide from the methanol. In doing so the methanol is regenerated and the acid gas for System 5, Sulfur Recovery, is produced.

The methanol containing the acid gas enters the Hot Regenerator after being heated in a shell and tube exchanger against hot regenerated methanol. In the Hot Regenerator, the dissolved gas is stripped by vapors produced in a steam-heated Reboiler plus methanol vapors from the methanol recovery subsystem.

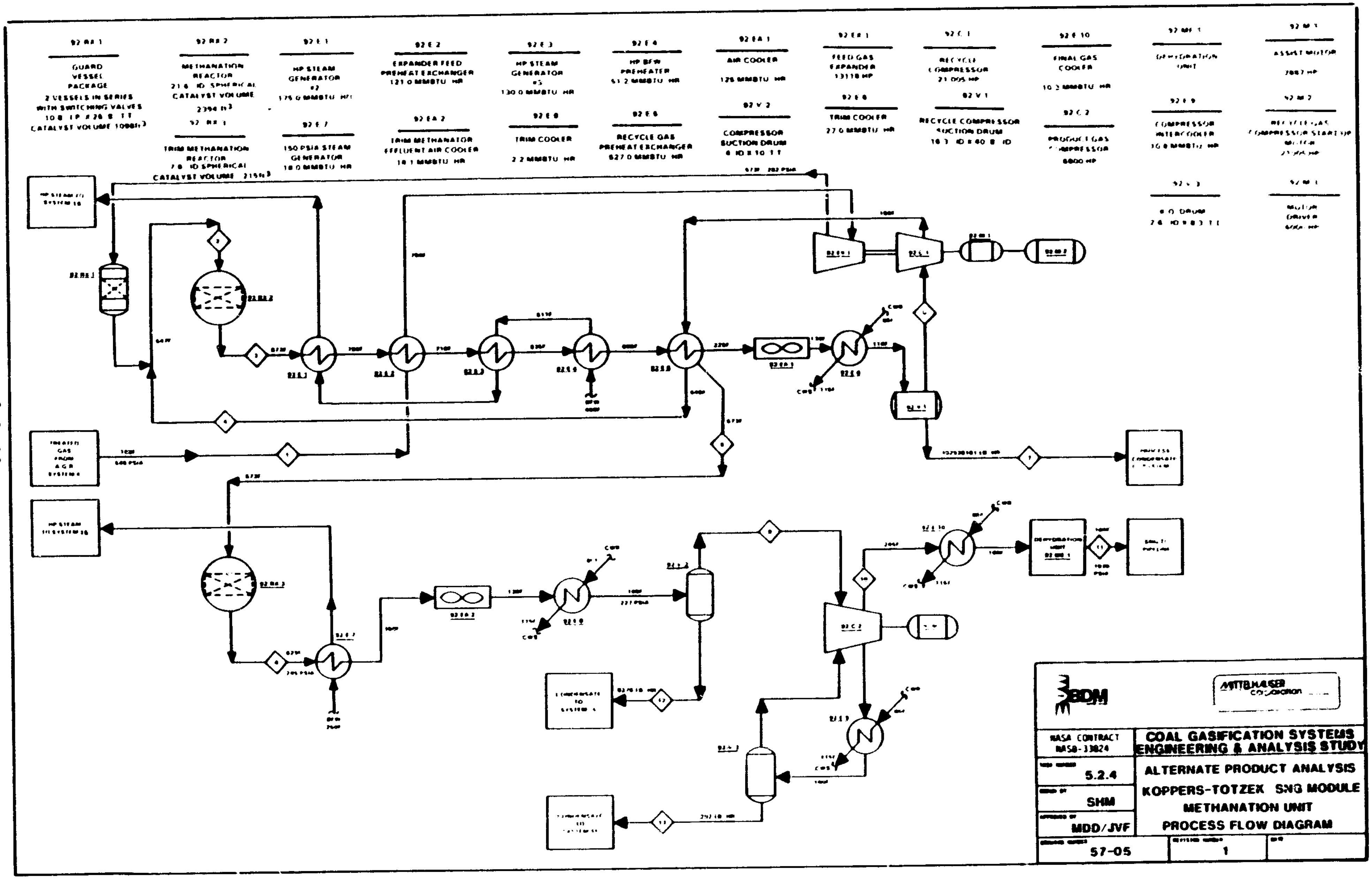
The methanol vapor leaving the top of the column is condensed out of the gas in a series of shell and tube exchangers. The condensed methanol accumulates in the reflux compartment of the tower and is pumped back to the top of the Hot Regenerator.

Part of the essentially gas free hot regenerated methanol leaving the bottom of the tower is used as a reflux in the methanol recovery subsystem. The remainder is exchanged with the hot regenerator feed and the interstage liquids in the Reabsorber and further pumped to the CO_2 Absorber.

System 20B: Methanation

Drawing 57-05 is a process flow diagram of this system. Table 4.4 contains the associated material balance. 26,089 moles per hour of treated gas from System 4, Acid Gas Removal, enters the methanation system at 100°F, 545 psia. The gas is first preheated to 750°F by exchange within 92-E-2, Expander Feed Preheat Exchanger hot methanator effluent. It is then let down to 282 psia in 92-EX-1, Feed Gas Expander, which generates about 13,100 HP. The expanded gas first passes through 92-V-1, Guard Vessel Package. This package consists of two vertical vessels each containing a bed of zinc oxide. The zinc oxide absorbs traces of sulfur compounds which remain in the fresh feed gas. The vessels are arranged in series, with piping and valving provided such that either vessel may be placed first in line, and such that one vessel may be taken offline, and the zinc oxide replaced while the other vessel is in service. The sulfur-free fresh feed is mixed with methanator recycle gas at a recycle/feed molar ratio of 5:1. The mixed gas temperature can be controlled by bypassing part of the gas around 92-E-5, Recycle Gas Preheat Exchanger, or by bypassing some of the fresh feed around the Feed Gas Expander.

The total feed gas, at a temperature of 547°F and an H_2/CO ratio of 3.34, then enters 92-V-2, Methanation Reactor. This vessel contains about 3000 cubic feet of a nickel catalyst which promotes the methanation of carbon oxides:



GUARD VESSEL PACKAGE
2 VESSELS IN SERIES WITH SWITCHING VALVES
METHANATION REACTOR
TRIM METHANATION REACTOR
HP STEAM GENERATOR
150 PSIA STEAM GENERATOR
EXPANDER FEED PREHEAT EXCHANGER
TRIM METHANATOR EFFLUENT AIR COOLER
HP STEAM GENERATOR
TRIM COOLER
HP BFW PREHEATER
RECYCLE GAS PREHEAT EXCHANGER
AIR COOLER
COMPRESSOR SUCTION DRUM
FEED GAS EXPANDER
TRIM COOLER
RECYCLE COMPRESSOR
RECYCLE COMPRESSOR SUCTION DRUM
FINAL GAS COOLER
PRODUCT GAS COMPRESSOR
DEHYDRATION UNIT
COMPRESSOR INTERCOOLER
ASSIST MOTOR
NASA CONTRACT NAS8-33824
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
ALTERNATE PRODUCT ANALYSIS
SHM
KOPPERS-TOTZEK SNG MODULE
METHANATION UNIT
MDD/JVF
PROCESS FLOW DIAGRAM
57-05
1
BDM
MITTELHAUSER CORPORATION

Table 4.4

Material Balance
Koppers-Totzek SNG Module
Methanation Unit

STREAM NUMBER			1		2		3	
STREAM ID			TREATED GAS FROM ACID GAS REMOVAL		METHANATION REACTOR FEED		METHANATION REACTOR OUTLET	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01	NOTE 1					
HYDROGEN	H_2	2.016	19,148.9		25,338.8		6,536.7	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	402.5		7,586.7		7,586.7	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	6,157.8		6,827.2		706.9	
CARBON DIOXIDE	CO_2	44.01	258.3		2,899.9		2,789.6	
METHANE	CH_4	16.042	119.8		113,466.2		119,696.8	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076	TR					
CARBONYL SULFIDE	COS	60.075	TR					
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			26,087.3		156,118.8		137,316.7	
WATER	H_2O	18.016	1.5		435.0		6,775.9	
TOTAL, WET			26,088.8	235,677	156,553.8	2,410,548	144,092.6	2,410,548
GAS MOLECULAR WEIGHT			9.03		15.40		17.55	
TEMPERATURE, °F			100		547		873	
PRESSURE, PSIA			640		282		277	
SCFH X 10^6			9.90		59.41		54.68	
GPM								

NOTE: THIS UNIT SIZED TO METHANATE 100% OF K-T MBG PRODUCT TO PROVIDE METHANATION UNIT CAPABLE OF HANDLING TEXACO MBG PRODUCT.

Table 4.4 (Continued)

Material Balance
Koppers-Totzek SNG Module
Methanation Unit

STREAM NUMBER			4		5		6	
STREAM ID			RECYCLE GAS TO METHANATION		METHANATION TRIM REACTOR FEED		RECYCLE GAS TO COMPRESSION	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR REACTOR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	6,189.9		346.8		6,189.9	
OXYGEN	O_2	12.0						
NITROGEN & ARGON	N_2	28.016	7,184.2		402.5		7,184.2	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	669.4		37.5		669.4	
CARBON DIOXIDE	CO_2	44.01	2,641.6		148.0		2,641.6	
METHANE	CH_4	16.042	113,346.4		6,350.4		113,346.4	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			130,031.5		7,285.8		130,031.5	
WATER	H_2O	18.016	433.5		365.8		6,410.1	
TOTAL, WET			130,465.0	2,174,871	7,651.0	128,000.4	136,441.6	2,282,545
GAS MOLECULAR WEIGHT			16.67		16.73		16.73	
TEMPERATURE, °F			150		573		110	
PRESSURE, PSIA			282		252		240	
SCFH X 10^6			49.51		2.9		51.78	
GPM								

Table 4.4 (Continued)

Material Balance
Koppers-Totzek SNG Module
Methanation Unit

STREAM NUMBER			7		8		9	
STREAM ID			METHANATION CONDENSATE TO TREATING		TRIM METHANATOR OUTLET		SNG TO PRODUCT COMPRESSOR	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016			64.9		64.9	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016			402.5		402.5	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01			1.8		1.8	
CARBON DIOXIDE	CO_2	44.01			104.3		104.3	
METHANE	CH_4	16.042			6,429.8		6,429.8	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY					7,003.3		7,003.3	
WATER	H_2O	18.016	5,976.6		488.9		28.6	
TOTAL, WET			5,976.6	107,674	7,492.2	128,000.4	7,031.9	119,544
GAS MOLECULAR WEIGHT			18		17.08		17.00	
TEMPERATURE, °F			110		629		100	
PRESSURE, PSIA			240		245		240	
SCFH X 10^6					2.85		2.67	
GPM			215					

Table 4.4 (Continued)

Material Balance
Koppers-Totzek SNG Module
Methanation Unit

STREAM NUMBER			10		11		12	
STREAM ID			SNG TO DEHYDRATION		SNG PRODUCT		TRIM METHANATION CONDENSATE TO TREATING	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	64.9		64.9			
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	402.5		402.5			
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	1.8		1.8			
CARBON DIOXIDE	CO_2	44.01	104.3		104.3			
METHANE	CH_4	16.042	6,429.8		6,429.8			
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	50.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			7,003.3		7,003.3			
WATER	H_2O	18.016	6.8		1.0		460.3	
TOTAL, WET			7,010.1	119.151	7,004.3	119.046	460.3	8,293
GAS MOLECULAR WEIGHT			17.0		17.0		18	
TEMPERATURE, °F			10		100		100	
PRESSURE, PSIA			1,030		1,015		240	
SCFH X 10^6			2.66		2.66			
GPM							17	

HHV = 929.9 BTU/SCF

Table 4.4 (Concluded)

Material Balance
Koppers-Totzek SNG Module
Methanation Unit

STREAM NUMBER			13					
STREAM ID			COMPRESSOR CONDENSATE TO TREATING					
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016						
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016						
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01						
CARBON DIOXIDE	CO_2	44.01						
METHANE	CH_4	16.042						
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO_2	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY								
WATER	H_2O	18.016	21.8					
TOTAL, WET			21.8					
GAS MOLECULAR WEIGHT			393					
TEMPERATURE, °F			18					
PRESSURE, PSIA			100					
SCFH			640					
GPM			1					

$$CO + 3H_2 \longrightarrow CH_4 + H_2O$$

$$CO_2 + 4H_2 \longrightarrow CH_4 + 2H_2O \quad .$$

The calculation of methanation yields is based on the reference design.[10] The operating conditions in the system were chosen to closely match those of the reference design.

The reactions shown above are exothermic, such that the effluent gas from the reactor is at 873°F. The gas is cooled in a series of shell and tube heat exchangers from 873°F to 110°F. Table 4.5 summarizes these exchangers.

Partially cooled net product gas is withdrawn from the cooling train at a temperature of 573°F. This temperature can be controlled by withdrawing part of the net product gas upstream of exchanger 92-E-5, and the balance downstream. This gas is sent to 92-V-4, Trim Methanation Reactor. The remaining recycle gas passes through 92-V-3, Compressor Suction Drum, where condensed water is separated from the recycle gas. The gas is compressed in 92-C-1, Recycle Compressor. The horsepower requirement in this compressor is approximately 21,000 BHP; about 62 percent of this is supplied by 92-EX-1, with the balance supplied by 92-M-1, Assist Motor. The Assist Motor is sized to supply the full power requirement of the compressor in case the expander must be removed from active service. This will permit continued operation of the methanation system between scheduled maintenance periods.

The net product gas, comprising 7,651 mol/hr still contains 4.53% H_2 and 0.5% CO; it is not suitable for pipeline gas. To convert the remaining H_2 and CO to CH_4, the gas is passed through 92-V-4, Trim Methanation Reactor. This spherical reactor is filled with 215 cubic feet of methanation catalyst, and converts 95.2 percent of the feed CO and about 29.5 percent of the feed CO_2 to CH_4. The effluent gas is cooled in 92-E-7, 150 psia Steam Generator; 92-EA-2, Trim Methanator Effluent Air Cooler; and 92-E-8, Trim Cooler from 629°F to 100°F. After

Table 4.5

Methanation Effluent Cooling Train

ITEM NUMBER	NAME	DUTY 10^6 BTU/HR	EFF. ΔT °F	U	AREA FT^2
E-1	HP STEAM GENERATOR #2	175.0	313.2	120	4,656
E-2	EXPANDER FEED PRE-HEATER EXCHANGER	121.0	192.5	60	10,476
E-3	HP STEAM GENERATOR #3	130	158.6	120	6,831
E-4	HP BFW PREHEATER	51.2	161.1	80	3,973
E-5	RECYCLE GAS PREHEAT EXCHANGER	572.0	67.2	60	130,704
EA-1	AIR COOLER	125.0	80.4	100	15,547
E-6	TRIM COOLER	27.0	10	80	33,750
E-7	150 PSIA STEAM GENERATOR	18.0	191.2	120	785
EA-2	TRIM METH. EFFLUENT AIR COOLER	18.1	136.4	100	1,327
E-8	TRIM COOLER	2.2	10	80	
HTR-1	HP STEAM SUPERHEATER	100.8			

separation of the condensate in 92-V-5, Compressor Suction Drum, the product gas is compressed in 92-C-2, Product Gas Compressor to about 1050 psia. This is a motor-driven centrifugal compressor with a power requirement of 6800 BHP. Finally, the product gas is cooled to 100°F, dehydrated to a maximum water content of 7 lbs per MMSCF with triethylene glycol in 92-ME-1, Dehydration Unit, and enters the pipeline at 1030 psia and 100°F.

The 63.84 MMSCFD of SNG from each K-T MBG upgrading plant module has a higher heating value of 930 BTU/SCF, a lifting index of 1.13, a flashback index of 1.03, a yellow-tip index of 1.19, and a CO content of 257 ppmv. All of these are acceptable values except the lifting index, which is high. This is due to the relatively high density of the gas caused by the pressure of N_2 and CO_2, without the instigating effect of H_2. Although the study budget did not permit repeating the design effort, a minor adjustment to the shift system operating conditions would allow additional hydrogen to leak through into the SNG, bringing the lifting index below its upper limit of 1.06. Such a minor adjustment would have a negligible impact on the cost of the product SNG.

Energy and Material Balance

Table 4.1 is the material balance for a K-T MBG upgrading plant module, while Table 4.6 is the energy balance. The efficiency of the module, expressed as HHV of SNG out divided by HHV of MBG in, is about 73.8%.

Operating Requirements

Table 4.7 shows the operating requirements for each system within the K-T MBG upgrading plant module. The totals are expressed as net impacts on the MBG facility.

Equipment List

The equipment list for the K-T MBG upgrading plant module is given in Appendix A.

Table 4.6

Energy Balance, 10^6 BTU Per Hour
Koppers-Totzek Process, System No.
Integrated SNG/Module Definition

ENTHALPY BALANCE

INLETS		PHASE	HHV	LATENT	SENSIBLE	TOTAL
STREAM	Coal	Solid	4,575		0	4,575
	Air	Gas		19.9	0	19.9
	Raw Water	Liquid			0	
SUBTOTAL IN						4,594.9
OUTLETS						
STREAM	SNG	Gas	2,222		0	2,222.0
	Ash	Solid				0
	Tail Gas	Gas		2.8	0.8	3.6
	Air Sep'n Vent	Gas		38.2	53.9	92.1
	Flue Gas	Gas		35.1	23.7	58.8
	Sulfur	Solid	61.5		0	61.5
SUBTOTAL OUT						2,438.0
NET, INLET - OUTLET						2,156.9
HEAT AND SHAFT WORK						
ELECTRIC POWER						425.8
SHAFT WORK						0
HEAT REJECTION				-828.2	-1,573.9	-2,402.1
HEAT INPUT (STEAM)						0
RADIATION/FUGITIVE LOSS						- 25
SUBTOTAL NET VALUE						-2,001.3
TOTAL ABSOLUTE VALUE						2,001.3

BALANCE: $\frac{\text{NET ENTHALPY + NET HEAT AND SHAFT WORK}}{\text{TOTAL ABSOLUTE VALUE OF HEAT AND SHAFT WORK}} \times 100 = +7.8$

BASIS: 60°F, LIQUID WATER

This balance represents one (1) Koppers MBG/SNG Module.

Table 4.7

Operating Requirements for Expected Operations
Koppers-Totzek Process
Integrated SNG/Module Definition

	BASIS	UNITS
Raw Materials		
Coal	TYP @ 100% operation	1,825,000 TYP
Raw Water	GYP @ 100% operation	1.156×10^9 gallons/year
Catalyst and Chemical Makeup		
Makeup	100 operation	$1,838,820/year
Initial Charge		$6,893,800
Utility Requirements		
Import Power	kWh/Yr @ 100% operation	1,092,8[illegible]508 kWhr/year
Operating Requirements		
Labor		
Supervisors	mh/year	29,120
Operators	mh/year	213,240
Supplies	Factored as 15 of Operating Labor Costs	
Maintenance Requirements		
Labor	Factored as 1.6 of Total Depreciable Direct Investment	
Supplies	Factored as 2.4 of Total Depreciable Direct Investment	

The above operating requirements represent one (1) Koppers MBG/SNG Module.

4.2 Texaco MBG Upgrading Module - Process Description

The MBG upgrading module consists of:

- System 20A: CO Shift
- System 4: Acid Gas Removal
- System 20B: Methanation.

Drawing 57-01 is a block flow diagram showing the arrangement of these systems within the module. Table 4.8 contains the associated material balance.

System 20A: CO Shift

Drawing 57-04 is a process flow diagram of this system. Table 4.9 contains the associated material balance. The feed gas consists of 32,818 moles per hour of raw gas from System 3, Gas Cooling, at 100°F and 690 psia. The configuration of the system is the same as that of the CO shift system in the K-T MBG upgrading plant. Table 4.10 summarizes the CO Shift Effluent Cooling Train.

If the soot scrubber in System 3 proved to be an effective particulate removal device, and if the NH_3 scrubber were not required, saturated gas at 455°F could be fed directly to the CO shift system, realizing a significant savings in steam, since the gas would contain essentially all of the water vapor required for the shift reaction, at the steam-to-dry gas ratios used in the Texaco design.

System 4: Acid Gas Removal

The design of the Acid Gas Removal System is identical to that in the K-T MBG upgrading plant module, except that the feed gas is 41,695 mols/hr instead of 40,314 moles/hr in the K-T case, and the CO_2 removal is 14,180 moles/hr instead of 13,470 moles/hr. However, the description for the K-T MBG upgrading plants' acid gas removal system will serve equally well for the Texaco MBG upgrading plan.

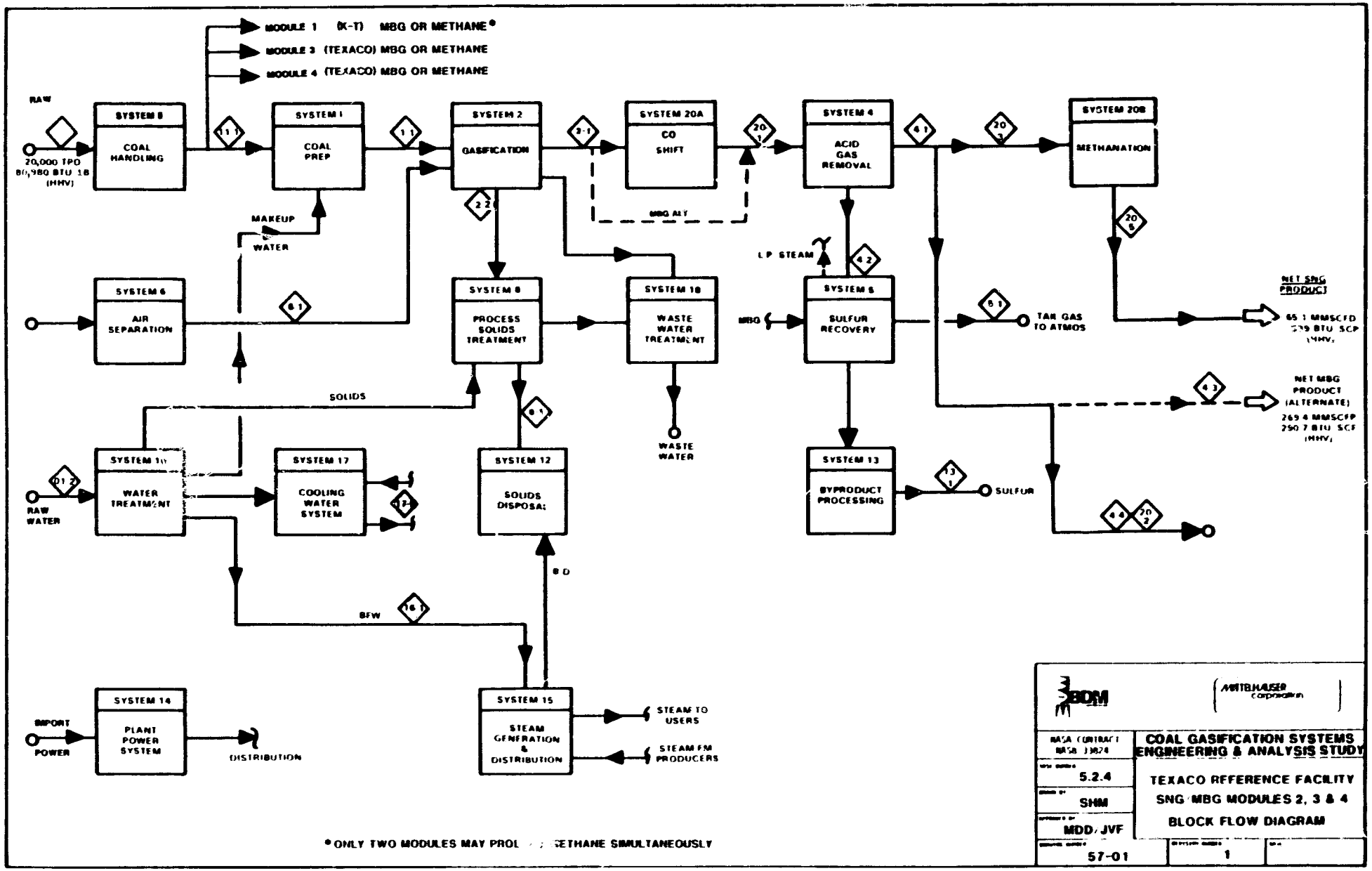

MODULE 1 (K-T) MBG OR METHANE*
MODULE 3 (TEXACO) MBG OR METHANE
MODULE 4 (TEXACO) MBG OR METHANE
RAW
20,000 TPD
8[illegible],980 BTU LB
(HHV)
SYSTEM 8
COAL HANDLING
SYSTEM 1
COAL PREP
SYSTEM 2
GASIFICATION
SYSTEM 20A
CO SHIFT
SYSTEM 4
ACID GAS REMOVAL
SYSTEM 20B
METHANATION
MAKEUP WATER
MBG ALT
LP STEAM
SYSTEM 6
AIR SEPARATION
SYSTEM 8
PROCESS SOLIDS TREATMENT
SYSTEM 18
WASTE WATER TREATMENT
MBG
SYSTEM 5
SULFUR RECOVERY
TAIL GAS TO ATMOS
NET SNG PRODUCT
65.1 MMSCFD
[illegible] BTU SCF
(HHV)
NET MBG PRODUCT (ALTERNATE)
269.4 MMSCFD
290.7 BTU SCF
(HHV)
SOLIDS
WASTE WATER
SYSTEM 16
WATER TREATMENT
RAW WATER
SYSTEM 17
COOLING WATER SYSTEM
SYSTEM 12
SOLIDS DISPOSAL
SYSTEM 13
BYPRODUCT PROCESSING
SULFUR
BFW
BD
SYSTEM 14
PLANT POWER SYSTEM
IMPORT POWER
DISTRIBUTION
SYSTEM 15
STEAM GENERATION & DISTRIBUTION
STEAM TO USERS
STEAM FM PRODUCERS
*ONLY TWO MODULES MAY PROD[illegible] METHANE SIMULTANEOUSLY
BOM
MITTELHAUSER Corporation
NASA CONTRACT NASA [illegible]
5.2.4
SHM
MDD/JVF
57-01
1
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
TEXACO REFERENCE FACILITY
SNG/MBG MODULES 2, 3 & 4
BLOCK FLOW DIAGRAM

Table 4.8

Material Balance
TEXACO Process
Integrated SNG/Module Definition

STREAM NUMBER			01-1*		01-2		1-1	
STREAM ID			RAW COAL FEED		RAW WATER MAKEUP		COAL SLURRY TO GASIFIER	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01		253,637				253,637
HYDROGEN	H_2	2.016		17,925				17,925
OXYGEN	O_2	32.0		23,901				23,901
NITROGEN & ARGON	N_2	28.016		5,761				5,761
SULFUR	S	32.06		15,449				15,449
CHLORINE	Cl	35.453		494				494
CARBON MONOXIDE	CO	28.01		-				-
CARBON DIOXIDE	CO_2	44.01		-				-
METHANE	CH_4	16.042		-				-
ETHYLENE	C_2H_4	28.052		-				-
ETHANE	C_2H_6	30.068		-				-
PROPYLENE	C_3H_6	42.078		-				-
PROPANE	C_3H_8	44.094		-				-
HYDROGEN SULFIDE	H_2S	34.076		-				-
CARBONYL SULFIDE	COS	60.075		-				-
CARBON DISULFIDE	CS_2	76.13		-				-
SULFUR DIOXIDE	SO_2	64.06		-				-
NITROUS OXIDE	NO	30.008		-				-
AMMONIA	NH_3	17.031		-				-
HYDROGEN CYANIDE	HCN	27.026		-				-
HYDROGEN CHLORIDE	HCl	36.461		-				-
NAPHTHA	-			-				-
TAR & OIL	-			-				-
ASH	-			59,650				59,650
OTHER SOLIDS				-				-
SUBTOTAL, DRY				376,817				376,817
WATER	H_2O	18.016		39,853				251,210
TOTAL, WET				416,670				628,027
GAS MOLECULAR WEIGHT			-				-	
TEMPERATURE, °F			-		66		140	
PRESSURE, PSIA			-		-		-	
SCFH			-		-		-	
GPM			-		2,290		-	

*MODULE 1 FRACTION ONLY.

Table 4.8 (Continued)

Material Balance
TEXACO Process
Integrated SNG/Module Definition

STREAM NUMBER			2-1		2-2		4-1	
STREAM ID			RAW GAS TO SHIFT		ASH FROM GASIFIER		SWEET GAS TO EXPANSION	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01	-	-			-	-
HYDROGEN	H_2	2.016	11,037.5	22,252			19,728.6	39,773
OXYGEN	O_2	32.0	-	-			-	-
NITROGEN & ARGON	N_2	28.016	375.	11,153			371.9	10,417
SULFUR	S	32.06	-	-			-	-
CHLORINE	Cl	35.453	-	-			-	-
CARBON MONOXIDE	CO	28.01	15,254.2	427,270			6,313.7	176,847
CARBON DIOXIDE	CO_2	44.01	5,466.7	240,589			144.1	6,342
METHANE	CH_4	16.042	154.2	2,474			147.8	2,371
ETHYLENE	C_2H_4	28.052	-	-			-	-
ETHANE	C_2H_6	30.068	-	-			-	-
PROPYLENE	C_3H_6	42.078	-	-			-	-
PROPANE	C_3H_8	44.094	-	-			-	-
HYDROGEN SULFIDE	H_2S	34.076	454.2	15,477			TR	TR
CARBONYL SULFIDE	COS	60.075	29.2	1,754			TR	TR
CARBON DISULFIDE	CS_2	76.13	-	-			-	-
SULFUR DIOXIDE	SO_2	64.06	-	-			-	-
NITROUS OXIDE	NO_2	30.008	-	-			-	-
AMMONIA	NH_3	17.031	104.2	1,775			-	-
HYDROGEN CYANIDE	HCN	27.026	-	-			-	-
HYDROGEN CHLORIDE	HCl	36.461	-	-			-	-
NAPHTHA	-		-	-			-	-
TAR & OIL	-		-	-			-	-
ASH	-		-	-		59,650	-	-
OTHER SOLIDS			-	-		3,053	-	-
SUBTOTAL, DRY			32,875.2	722,744		62,703	26,706.1	235,749
WATER	H_2O	18.016	66,543.9	1,198,855		-	1.5	27
TOTAL, WET			99,419.1	1,921,599		62,703	26,707.6	235,776
GAS MOLECULAR WEIGHT			19.33		-		8.8	
TEMPERATURE, °F			455		-		100	
PRESSURE, PSIA			690		-		620	
SCFH			37,729,548		-		10,140,019	
GPM			-		-		-	

Table 4.8 (Continued)

Material Balance
TEXACO Process
Integrated SNG/Module Definition

STREAM NUMBER			4-2		4-3		4-4	
STREAM ID			ACID GAS TO SULFUR RECOVERY		NET MBG PRODUCT (100% MBG PRODUCTION)		MBG INTERNAL USAGE (100% MBG PRODUCTION)	
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR
CARBON	C	12.01	-	-	-	-	-	-
HYDROGEN	H_2	2.016	-	-	10,965.8	22,107	67.2	135
OXYGEN	O_2	32.0	-	-	-	-	-	-
NITROGEN & ARGON	N_2	28.016	0.1	2.8	372.7	11,085	2.3	68
SULFUR	S	32.06	-	-	-	-	-	-
CHLORINE	Cl	35.453	-	-	-	-	-	-
CARBON MONOXIDE	CO	28.01	-	-	15,157.8	424,567	92.9	2,602
CARBON DIOXIDE	CO_2	44.01	2,678.5	75,025	2,919.6	128,492	17.9	789
METHANE	CH_4	16.042	-	-	153.3	2,459	0.9	15.1
ETHYLENE	C_2H_4	28.052	-	-	-	-	-	-
ETHANE	C_2H_6	30.068	-	-	-	-	-	-
PROPYLENE	C_3H_6	42.078	-	-	-	-	-	-
PROPANE	C_3H_8	44.094	-	-	-	-	-	-
HYDROGEN SULFIDE	H_2S	34.076	479.5	16,341	TR	TR	TR	TR
CARBONYL SULFIDE	COS	60.075	3.3	198	TR	TR	TR	TR
CARBON DISULFIDE	CS_2	76.13	-	-	-	-	-	-
SULFUR DIOXIDE	SO_2	64.06	-	-	-	-	-	-
NITROUS OXIDE	NO	30.008	-	-	-	-	-	-
AMMONIA	NH_3	17.031	-	-	-	-	-	-
HYDROGEN CYANIDE	HCN	27.026	-	-	-	-	-	-
HYDROGEN CHLORIDE	HCl	36.461	-	-	-	-	-	-
NAPHTHA	-		-	-	-	-	-	-
TAR & OIL	-		-	-	-	-	-	-
ASH	-		-	-	-	-	-	-
OTHER SOLIDS			-	-	-	-	-	-
SUBTOTAL, DRY			3,164.4	134,423	29,575.1	589,035	181.2	3,610
WATER	H_2O	18.016	50	901	4.4	78	0.1	1.2
TOTAL, WET			3,311.4	135,324	29,579.4	589,114	181.3	3,611
GAS MOLECULAR WEIGHT			42.1		19.92		19.92	
TEMPERATURE, °F			120		60		57	
PRESSURE, PSIA			22		615		615	
SCFH			1,219,266		11,225,403		68,812	
GPM			-		-		-	

Table 4.8 (Continued)

Material Balance
TEXACO Process
Integrated SNG/Module Definition

STREAM NUMBER			5-1		6-1		8-1	
STREAM ID			TAIL GAS FROM SULFUR RECOVERY		OXIDANT TO GASIFIER		RPOCESS SOLIDS TO DISPOSAL	
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR	LB-MOLS HR	LBS HR
CARBON	C	12.01	-	-	-	-		
HYDROGEN	H_2	2.016	25.8	52	-	-		
OXYGEN	O_2	32.0	-	-	-	-		
NITROGEN & ARGON	N_2	28.016	1,000.7	28,036	10,787.9	345,212		
SULFUR	S	32.06	-	-	-	-		
CHLORINE	Cl	35.453	-	-	-	-		
CARBON MONOXIDE	CO	28.01	-	-	-	-		
CARBON DIOXIDE	CO_2	44.01	2,587	113,871	-	-		
METHANE	CH_4	16.042	-	-	-	-		
ETHYLENE	C_2H_4	28.052	-	-	-	-		
ETHANE	C_2H_6	30.068	-	-	-	-		
PROPYLENE	C_3H_6	42.078	-	-	-	-		
PROPANE	C_3H_8	44.094	-	-	-	-		
HYDROGEN SULFIDE	H_2S	34.076	TR	TR	-	-		
CARBONYL SULFIDE	COS	60.075	TR	TR	-	-		
CARBON DISULFIDE	CS_2	76.13	-	-	-	-		
SULFUR DIOXIDE	SO_2	64.06	-	-	-	-		
NITROUS OXIDE	NO	30.008	-	-	-	-		
AMMONIA	NH_3	17.031	-	-	-	-		
HYDROGEN CYANIDE	HCN	27.026	-	-	-	-		
HYDROGEN CHLORIDE	HCl	36.461	-	-	-	-		
NAPHTHA	-		-	-	-	-		
TAR & OIL	-		-	-	-	-		
ASH	-		-	-	-	-		59,650
OTHER SOLIDS			-	-	-	-		3,203
SUBTOTAL, DRY			3,613.9	141,959	11,008.1	352,037		62,853
WATER	H_2O	18.016	167.9	3,025	-	-		150
TOTAL, WET			3,781.8	144,984	11,008.1	352,037		63,003
GAS MOLECULAR WEIGHT			38.3		31.98			
TEMPERATURE, °F			100		300			
PRESSURE, PSIA			14.4		815			
SCFH			1,435,193		4,177,574			
GPM			-		-			

Table 4.8 (Continued)

Material Balance
TEXACO Process
Integrated SNG/Module Definition

STREAM NUMBER			11-1		13-1		16-1	
STREAM ID			COAL TO COAL PREPARATION		SULFUR BYPRODUCT		MAKEUP WATER TO BDM SYSTEM	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01		253,637	-	-		
HYDROGEN	H_2	2.016		17,925	-	-		
OXYGEN	O_2	32.0		23,501	-	-		
NITROGEN & ARGON	N_2	28.016		5,761	-	-		
SULFUR	S	32.06		15,449	482.8	15,478		
CHLORINE	Cl	35.453		494	-	-		
CARBON MONOXIDE	CO	28.01		-	-	-		
CARBON DIOXIDE	CO_2	44.01		-	-	-		
METHANE	CH_4	16.042		-	-	-		
ETHYLENE	C_2H_4	28.052		-	-	-		
ETHANE	C_2H_6	30.068		-	-	-		
PROPYLENE	C_3H_6	42.078		-	-	-		
PROPANE	C_3H_8	44.094		-	-	-		
HYDROGEN SULFIDE	H_2S	34.076		-	-	-		
CARBONYL SULFIDE	COS	60.075		-	-	-		
CARBON DISULFIDE	CS_2	76.13		-	-	-		
SULFUR DIOXIDE	SO_2	64.06		-	-	-		
NITROUS OXIDE	NO	30.008		-	-	-		
AMMONIA	NH_3	17.031		-	-	-		
HYDROGEN CYANIDE	HCN	27.026		-	-	-		
HYDROGEN CHLORIDE	HCl	36.461		-	-	-		
NAPHTHA	-			-	-	-		
TAR & OIL	-			-	-	-		
ASH	-			59,650	-	-		
OTHER SOLIDS				-	-	-		
SUBTOTAL, DRY				376,817	482.8	15,478		
WATER	H_2O	18.016		39,853	-	-		
TOTAL, WET				416,670	482.8	15,478		
GAS MOLECULAR WEIGHT			-		-			
TEMPERATURE, °F			-		60			
PRESSURE, PSIA					-			
SCFH			-		-			
GPM			-		-		85.4	

Table 4.8 (Continued)

Material Balance
TEXACO Process
Integrated SNG/Module Definition

STREAM NUMBER			20-1		20-2		20-3	
STREAM ID			SHIFTED GAS TO ACID GAS REMOVAL		MBG INTERNAL USAGE (100% SNG PRODUCTION)		EXPANDED SHIFTED GAS TO METHANATION (100% SNG PRODUCTION)	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01	-	-	-	-	-	-
HYDROGEN	H_2	2.016	19,837.8	40,065	120.8	243	19,607.8	39,529
OXYGEN	O_2	32.0	-	-	-	-	-	-
NITROGEN & ARGON	N_2	28.016	375.1	10,506	2.3	64	369.6	10,352
SULFUR	S	32.06	-	-	-	-	-	-
CHLORINE	Cl	35.453	-	-	-	-	-	-
CARBON MONOXIDE	CO	28.01	6,417.9	179,765	38.7	1,084	6,275.0	175,763
CARBON DIOXIDE	CO_2	44.01	14,328.3	630,588	0.9	40	143.2	6,302
METHANE	CH_4	16.042	154.3	2,475	0.9	14	146.9	2,356
ETHYLENE	C_2H_4	28.052	-	-	-	-	-	-
ETHANE	C_2H_6	30.068	-	-	-	-	-	-
PROPYLENE	C_3H_6	42.078	-	-	-	-	-	-
PROPANE	C_3H_8	44.094	-	-	-	-	-	-
HYDROGEN SULFIDE	H_2S	34.076	479.6	16,345	TR	TR	TR	TR
CARBONYL SULFIDE	COS	60.075	3.9	234	TR	TR	TR	TR
CARBON DISULFIDE	CS_2	76.13	-	-	-	-	-	-
SULFUR DIOXIDE	SO_2	64.06	-	-	-	-	-	-
NITROUS OXIDE	NO	30.008	-	-	-	-	-	-
AMMONIA	NH_3	17.031	-	-	-	-	-	-
HYDROGEN CYANIDE	HCN	27.026	-	-	-	-	-	-
HYDROGEN CHLORIDE	HCl	36.461	-	-	-	-	-	-
NAPHTHA	-		-	-	-	-	-	-
TAR & OIL	-		-	-	-	-	-	-
ASH	-		-	-	-	-	-	-
OTHER SOLIDS			-	-	-	-	-	-
SUBTOTAL, DRY			41,632.9	879,978	163.59	1,439.5	26,542.6	234,303
WATER	H_2O	18.016	62	1,117	0.01	0.2	1.39	25
TOTAL, WET			41,694.9	881,095	163.6	1,439.7	26,544	234,328
GAS MOLECULAR WEIGHT			21.1		8.8		8.8	
TEMPERATURE, °F			100		100		100	
PRESSURE, PSIA			640		297		297	
SCFH			15,830,217		63,168		10,077,906	
GPM			-		-		-	

Table 4.8 (Concluded)

Material Balance
TEXACO Process
Integrated SNG/Module Definition

STREAM NUMBER			20-4		20-5 NET SNG		◇	
STREAM ID			PRODUCT SNG TO COMPRESSION		PRODUCT (100% SNG PRODUCTION)			
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01	-	-	-	-		
HYDROGEN	H_2	2.016	212.9	429	212.9	429		
OXYGEN	O_2	32.0	-	-	-	-		
NITROGEN & ARGON	N_2	28.016	369.6	10,352	369.6	10,352		
SULFUR	S	32.06	-	-	-	-		
CHLORINE	Cl	35.453	-	-	-	-		
CARBON MONOXIDE	CO	28.01	0.04	1.1	0.04	1.1		
CARBON DIOXIDE	CO_2	44.01	0.75	33	0.75	33		
METHANE	CH_4	16.042	6,564.4	105,293	6,564.4	105,293		
ETHYLENE	C_2H_4	28.052	-	-	-	-		
ETHANE	C_2H_6	30.068	-	-	-	-		
PROPYLENE	C_3H_6	42.078	-	-	-	-		
PROPANE	C_3H_8	44.094	-	-	-	-		
HYDROGEN SULFIDE	H_2S	34.076	TR	TR	TR	TR		
CARBONYL SULFIDE	COS	60.075	TR	TR	TR	TR		
CARBON DISULFIDE	CS_2	76.13	-	-	-	-		
SULFUR DIOXIDE	SO_2	64.06	-	-	-	-		
NITROUS OXIDE	NO	30.008	-	-	-	-		
AMMONIA	NH_3	17.031	-	-	-	-		
HYDROGEN CYANIDE	HCN	27.026	-	-	-	-		
HYDROGEN CHLORIDE	HCl	36.461	-	-	-	-		
NAPHTHA	-		-	-	-	-		
TAR & OIL	-		-	-	-	-		
ASH	-		-	-	-	-		
OTHER SOLIDS			-	-	-	-		
SUBTOTAL, DRY			7,147.7	116,108	7,147.7	116,109		
WATER	H_2O	18.016	31.1	560	1.06	19		
TOTAL, WET			7,178.8	116,669	7,148.8	116,128		
GAS MOLECULAR WEIGHT			16.2		16.2			
TEMPERATURE, °F			100		60			
PRESSURE, PSIA			227		1030			
SCFH			2,725,560		2,714,170			
GPM			-		-			

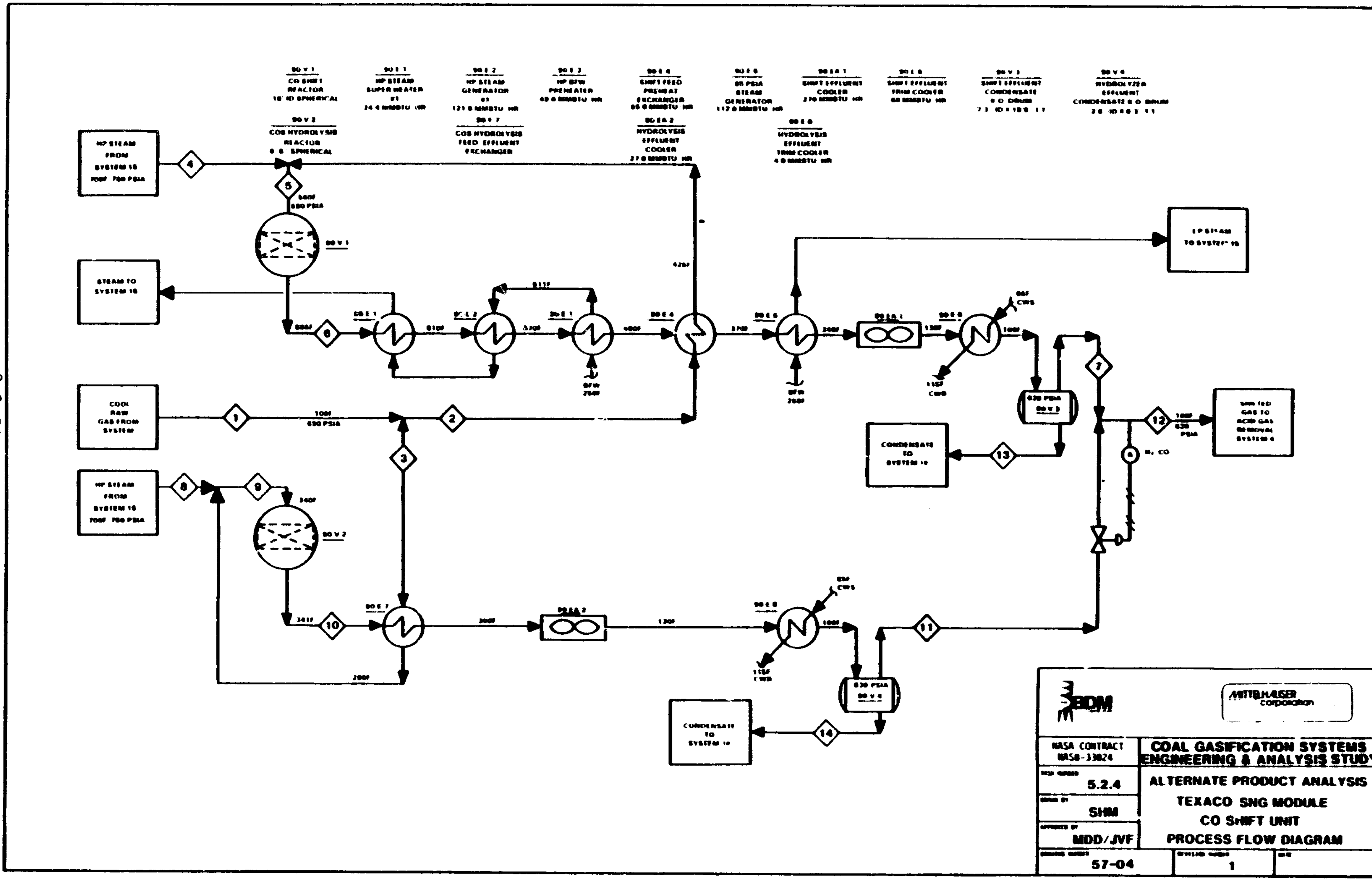

HP STEAM FROM SYSTEM 15 700F 750 PSIA
STEAM TO SYSTEM 15
COOL RAW GAS FROM SYSTEM
HP STEAM FROM SYSTEM 15 700F 750 PSIA
100F 690 PSIA
LP STEAM TO SYSTEM 15
CONDENSATE TO SYSTEM 10
SHIFTED GAS TO ACID GAS REMOVAL SYSTEM 4
CONDENSATE TO SYSTEM 10
BDM
MITTELHAUSER corporation
NASA CONTRACT NAS8-33824
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
ALTERNATE PRODUCT ANALYSIS
SHM
TEXACO SNG MODULE
CO SHIFT UNIT
MDD/JVF
PROCESS FLOW DIAGRAM
57-04
1

Table 4.9

Material Balance
TEXACO SNG Module
CO Shift Unit

STREAM NUMBER			1		2		3	
STREAM ID			GASIFIER EFFLUENT TO SHIFT		CO SHIFT REACTOR FEED		COS HYDROLYSIS REACTOR FEED	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.1						
HYDROGEN	H_2	2.016	11,037.5		8,278.1		2,759.4	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	375.0		281.3		93.8	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	15,254.2		11,440.6		3,813.6	
CARBON DIOXIDE	CO_2	44.01	5,466.7		4,100.0		1,373.8	
METHANE	CH_4	16.042	154.2		115.7		93.8	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076	454.2		340.7		120.7	
CARBONYL SULFIDE	COS	60.075	29.2		21.9		0.2	
CARBON DISULFIDE	CS	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO_2	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			32,771.0		24,578.3		8,193.0	
WATER	H_2O	18.016	47.5		35.6		11.9	
TOTAL, WET			32,818.5	721,825	24,613.9	559,636	8,204.9	162,189
GAS MOLECULAR WEIGHT			22.0		22.74		19.77	
TEMPERATURE, °F			100		100		100	
PRESSURE, PSIA			690		690		690	
SCFH X 10^6			12.45		9.34		3.11	
GPM	-							

Table 4.9 (Continued)

Material Balance
TEXACO SNG Module
CO Shift Unit

STREAM NUMBER			4		5		6	
STREAM ID			HP STEAM TO CO SHIFT		CO SHIFT REACTOR INLET		CO SHIFT REACTOR OUTLET	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016			8,278.1		17,114.4	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016			281.3		281.3	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01			11,440.6		2,604.3	
CARBON DIOXIDE	CO_2	44.01			4,100.0		12,954.5	
METHANE	CH_4	16.042			115.7		115.7	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076			340.7		358.9	
CARBONYL SULFIDE	COS	60.075			21.9		3.7	
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS	-							
SUBTOTAL, DRY					24,578.3		33,432.8	
WATER	H_2O	18.016	26,564.4		26,600.4		17,745.9	
TOTAL, WET			26,564.4	478,585	51,178.7	1,038,221	51,178.7	1,038,221
GAS MOLECULAR WEIGHT			18		20.29		20.29	
TEMPERATURE, °F			700		560		886	
PRESSURE, PSIA			750		680		655	
SCFH X 10^6			10.08		19.42		19.42	
GPM								

Table 4.9 (Continued)

Material Balance
TEXACO SNG Module
CO Shift Unit

STREAM NUMBER			7		8		9	
STREAM ID			SHIFT EFFLUENT		HP STEAM TO COS HYDROLYSIS REACTOR		COS HYDROLYSIS REACTOR INLET	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01					2,759.4	
HYDROGEN	H_2	2.016	17,114.4					
OXYGEN	O_2	32.0					93.8	
NITROGEN & ARGON	N_2	28.016	281.3					
SULFUR	S	32.06						
CHLORINE	Cl	35.453					3,813.6	
CARBON MONOXIDE	CO	28.01	2,604.3				1,373.8	
CARBON DIOXIDE	CO_2	44.01	12,954.5				93.8	
METHANE	CH_4	16.042	115.7					
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094					120.7	
HYDROGEN SULFIDE	H_2S	34.076	358.9				0.2	
CARBONYL SULFIDE	COS	60.07[illegible]	3.7					
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS							8,193.0	
SUBTOTAL, DRY			33,432.8				1,445.6	
WATER	H_2O	18.016	54.0		1,433.7		9,638.6	188,019
TOTAL, WET			33,486.8	719,484	1,433.7	25,330	19.51	
GAS MOLECULAR WEIGHT			21.49		18.0		340	
TEMPERATURE, °F			100		700		680	
PRESSURE, PSIA			620		750		3.66	
SCFH X 10^6			12.71		0.5			
GPM								

Table 4.9 (Continued)

Material Balance
TEXACO SNG Module
CO Shift Unit

STREAM NUMBER			10		11		12	
STREAM ID			COS HYDROLYSIS REACTOR OUTLET		COS HYDROLYSIS REACTOR EFFLUENT		SHIFTED GAS TO ACID GAS REMOVAL	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	2,759.4		2,759.4		19,873.8	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	93.8		93.8		375.1	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	3,813.6		3,813.6		6,417.9	
CARBON DIOXIDE	CO_2	44.01	1,373.8		1,373.8		14,328.3	
METHANE	CH_4	16.042	38.6		38.6		154.3	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076	120.7		120.7		479.6	
CARBONYL SULFIDE	COS	60.075	0.2		0.2		3.9	
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			8,200.1		8,200.1		41,632.9	
WATER	H_2O	18.016	1,438.5		13.0		67.0	
TOTAL, WET			9,638.6	188,019	8,213.1	162,337	41,699.9	881,821
GAS MOLECULAR WEIGHT			19.51		19.77		21.15	
TEMPERATURE, °F			341		100		100	
PRESSURE, PSIA			655		630		615	
SCFH X 10^6			3.56		3.11		15.83	
GPM								

Table 4.9 (Concluded)

Material Balance
TEXACO SNG Module
CO Shift Unit

STREAM NUMBER			13		14		◇	
STREAM ID			SHIFT CONDENSATE TO TREATING		COS HYDROLYSIS CONDENSATE TO TREATING			
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016						
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016						
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01						
CARBON DIOXIDE	CO_2	44.01						
METHANE	CH_4	16.042						
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY								
WATER	H_2O	18.016	17,691.9		1,425.5			
TOTAL, WET			17,691.9	318,737	1,425.5	25,682		
GAS MOLECULAR WEIGHT			18.0		18.0			
TEMPERATURE, °F			100		100			
PRESSURE, PSIA			620		630			
SCFH								
GPM			638		52			

Table 4.10

CO SHIFT EFFLUENT COOLING TRAIN - TEXACO

ITEM NUMBER	NAME	DUTY 10^6 BTU/HR	EFF. ΔT °F	U	AREA FT^2
E-1	HP STEAM SUPERHEATER #1	24.4	238.1	60	1,200
E-2	HP STEAM GENERATOR #1	121.6	147.9	120	6,852
E-3	HP BFW PRE-HEATER	49.0	125.7	80	4,873 (2 Shells)
E-4	SHIFT FEED PREHEAT EXCHANGER	65.0	134.2	60	8,072
E-5	65 PSIA STEAM GENERATOR	112.8	80.7	120	11,655
EA-1	SHIFT EFFLUENT COOLER	270.0	123.1	100	21,925
E-6	SHIFT EFFLUENT TRIM COOLER	160.0	10	80	75,000 (2 shells)
E-7	COS HYDROLYSIS FEED/EFFLUENT EXCHANGER	12.2	117.1	60	1,740
EA-2	HYDROLYSIS EFFLUENT COOLER	27.8	109.2	100	2,545
E-8	HYDROLYSIS EFFLUENT TRIM COOLER	4.8	10	80	6,000 (3 shells)

System 20B: Methanation

Drawing 57-03 is a process flow diagram of the methanation system. Table 4.11 contains the associated material balance. As with the CO shift system, the process configuration and description are the same as in the K-T MBG upgrading plant; the Texaco module is somewhat larger because the gas feed to the methanation system is 13.7% greater than in the K-T case. The product SNG has a higher heating value of 939.0 BTU/SCF, a lifting index of 1.07, a flashback index of 1.038, a yellow-tip index of 1.141, and a CO content of about 5 ppmv. This means that the SNG product is essentially compatible with natural gas. Table 4.12 summarizes the Methanation Effluent Cooling Train.

Energy and Material Balance

Tables 4.8 and 4.9 are the material balance for a Texaco MBG upgrading plant module, while Table 4.13 is the energy balance. The energy balance shows that the efficiency of the module, expressed as HHV of SNG out divided by HHV of MBG in, is about 74.5%.

Operating Requirements

Table 4.14 shows the operating requirements for each system within the Texaco MBG upgrading plant module. The totals are expressed as net impacts on the MBG facility.

Equipment List

The equipment list for the Texaco MBG upgrading plant module is given in Appendix B.

4.3 Utility System Impacts

Koppers-Totzek MBG Upgrading Module

Design No. 1 consists of two K-T MBG upgrading plant modules, serving an MBG facility consisting of four parallel K-T MBG modules. Drawing 57-07 is a diagram of System 15, Steam Generation/Distribution,

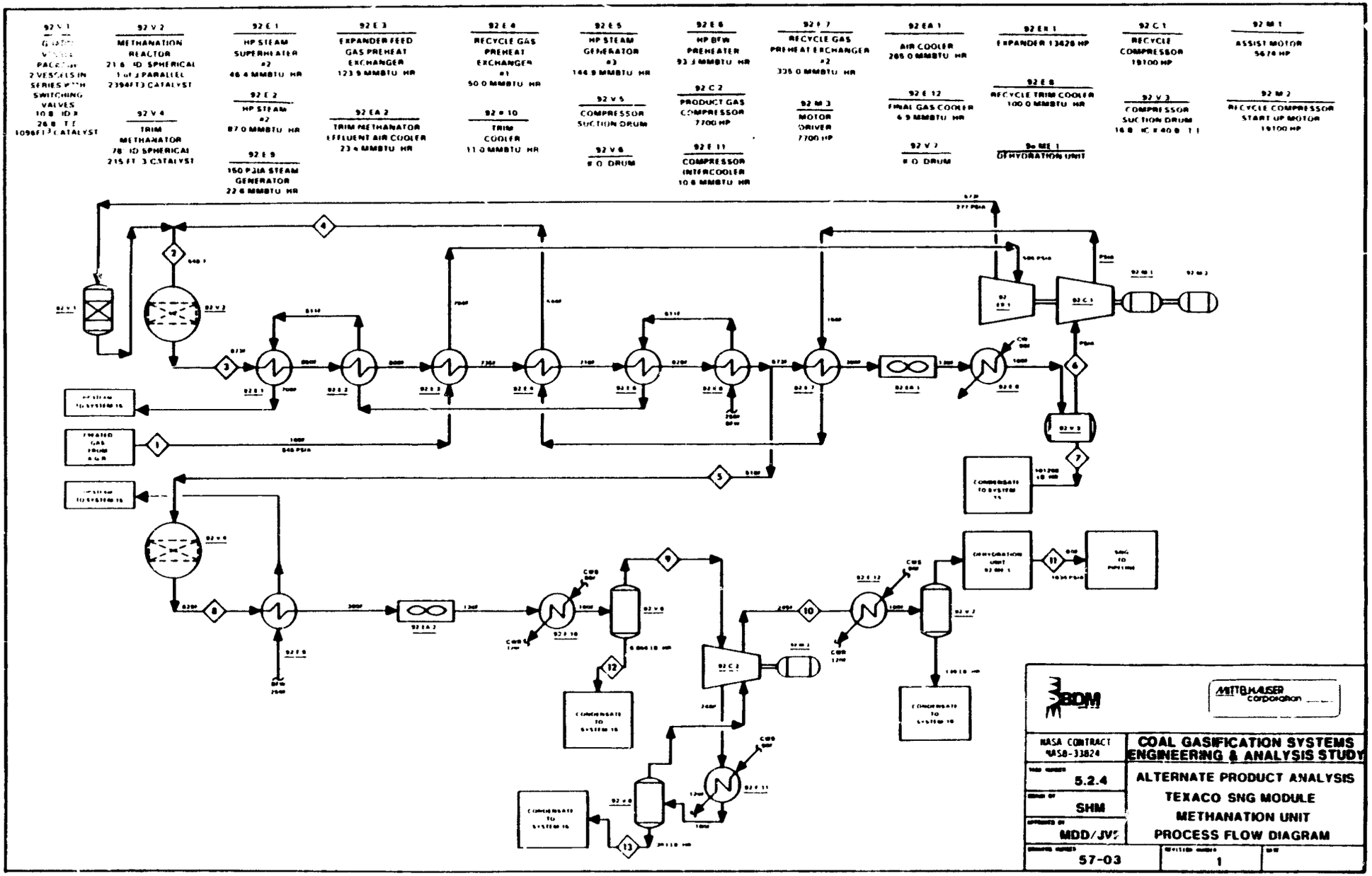

92 V 1
92 V 2
METHANATION REACTOR
92 V 4
TRIM METHANATOR
92 E 1
HP STEAM SUPERHEATER #2
92 E 2
HP STEAM #2
92 E 9
150 PSIA STEAM GENERATOR
92 E 3
EXPANDER FEED GAS PREHEAT EXCHANGER
92 EA 2
TRIM METHANATOR EFFLUENT AIR COOLER
92 E 4
RECYCLE GAS PREHEAT EXCHANGER #1
92 # 10
TRIM COOLER
92 E 5
HP STEAM GENERATOR #3
92 V 5
COMPRESSOR SUCTION DRUM
92 V 6
92 E 6
HP BFW PREHEATER
92 C 2
PRODUCT GAS COMPRESSOR
7700 HP
92 E 11
COMPRESSOR INTERCOOLER
92 E 7
RECYCLE GAS PREHEAT ERCHANGER #2
92 M 3
MOTOR DRIVER
7700 HP
92 EA 1
AIR COOLER
92 E 12
FINAL GAS COOLER
92 V 7
92 E R 1
EXPANDER 13426 HP
92 E 8
RECYCLE TRIM COOLER
92 ME 1
DEHYDRATION UNIT
92 C 1
RECYCLE COMPRESSOR
19100 HP
92 V 3
COMPRESSOR SUCTION DRUM
92 M 1
ASSIST MOTOR
5670 HP
92 M 2
RECYCLE COMPRESSOR START UP MOTOR
19100 HP
MITTELHAUSER corporation
NASA CONTRACT NAS8-33824
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
ALTERNATE PRODUCT ANALYSIS
TEXACO SNG MODULE
METHANATION UNIT
PROCESS FLOW DIAGRAM
SHM
MDD/JVS
57-03
1

Table 4.11

Material Balance
TEXACO SNG Module
Methanation Unit

STREAM NUMBER			1		2		3	
STREAM ID			TREATED GAS FROM ACID GAS REMOVAL		METHANATION REACTOR FEED		METHANATION REACTOR OUTLET	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	19,728.6		27,720.6		8,436.0	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	371.9		7,066.1		7,066.1	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	6,313.7		6,641.3		345.8	
CARBON DIOXIDE	CO_2	44.01	144.7		946.9		847.4	
METHANE	CH_4	16.042	147.8		117,918.2		124,313.2	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076	TR					
CARBONYL SULFIDE	COS	60.075	TR					
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			26,706.1		160,293.1		141,008.5	
WATER	H_2O	18.016	1.5		224.1		6,718.6	
TOTAL, WET			26,707.6	235,776	160,517.2	2,377,226	147,727.1	2,337,226
GAS MOLECULAR WEIGHT			8.8		14.81		16.09	
TEMPERATURE, °F			100		546		873	
PRESSURE, PSIA			620		282		277	
SCFH X 10^6			10.14		60.92		56.06	
GPM								

Table 4.11 (Continued)

Material Balance
TEXACO SNG Module
Methanation Unit

STREAM NUMBER			4		5		6	
STREAM ID			RECYCLE GAS TO METHANATION REACTOR		TRIM METHANATION REACTOR FEED		RECYCLE GAS TO COMPRESSION	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	7,992.0		444.0		7,992.0	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	6,694.2		371.9		6,694.2	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	327.6		18.2		327.6	
CARBON DIOXIDE	CO_2	44.01	802.8		44.6		802.8	
METHANE	CH_4	16.042	117,770.4		6,546.8		117,770.4	
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			133,587.0		7,421.5		133,587.0	
WATER	H_2O	18.016	222.6		353.6		222.6	
TOTAL, WET			133,809.6	2,141,450	7,775.1	125,117	133,809.6	2,141,450
GAS MOLECULAR WEIGHT			16.00		16.09		16.00	
TEMPERATURE, °F			540		510		100	
PRESSURE, PSIA			282		292		277	
SCFH X 10^6			50.78		2.95		50.78	
GPM								

Table 4.11 (Continued)

Material Balance
TEXACO SNG Module
Methanation Unit

STREAM NUMBER			7		8		9	
STREAM ID			METHANATION CONDENSATE TO TREATING		TRIM METHANATION REACTOR OUTLET		SNG TO PRODUCT COMPRESSOR	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016			214.2		214.2	
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016			371.9		371.9	
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01			TR		TR	
CARBON DIOXIDE	CO_2	44.01			0.8		0.8	
METHANE	CH_4	16.042			6,604.8		6,604.8	
ETHYLENE	C_2H_4	26.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY					7,191.7		7,191.7	
WATER	H_2O	18.016	6,412.8		459.4		26.1	
TOTAL, WET			6,412.6	110,658	7,651.1	127,023	7,217.8	119,217
GAS MOLECULAR WEIGHT					16.6		16.52	
TEMPERATURE, °F			100		282		100	
PRESSURE, PSIA			240		629		277	
SCFH X 10^6					2.90		2.74	
GPM								

Table 4.11 (Continued)

Material Balance
TEXACO SNG Module
Methanation Unit

STREAM NUMBER			10		11		12	
STREAM ID			SNG TO DEHYDRATION		SNG PRODUCT		TRIM METHANATION CONDENSATE TO	
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS TREATING/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016	214.2		214.2			
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016	317.9		317.9			
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01	TR		TR			
CARBON DIOXIDE	CO_2	44.01	0.8		0.8			
METHANE	CH_4	16.042	6,604.8		6,604.8			
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO_2	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY			7,191.7		7,191.7			
WATER	H_2O	18.016	7.0		1.0		433.3	
TOTAL, WET			7,191.8	118,873	7,192.7	118,764	433.3	7,806
GAS MOLECULAR WEIGHT			16.53		16.51			
TEMPERATURE, °F			100		60		100	
PRESSURE, PSIA			1,030		1,015		200	
SCFH X 10^6			2.73		2.73			
GPM							1(	

HHV = 939.0 BTU/SCF

Table 4.11 (Continued)

Material Balance
TEXACO SNG Module
Methanation Unit

STREAM NUMBER			13					
STREAM ID			COMPRESSOR CONDENSATE TO TREATING					
	SYMBOL	FORMULA WEIGHT	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR	LB-MOLS/HR	LBS/HR
CARBON	C	12.01						
HYDROGEN	H_2	2.016						
OXYGEN	O_2	32.0						
NITROGEN & ARGON	N_2	28.016						
SULFUR	S	32.06						
CHLORINE	Cl	35.453						
CARBON MONOXIDE	CO	28.01						
CARBON DIOXIDE	CO_2	44.01						
METHANE	CH_4	16.042						
ETHYLENE	C_2H_4	28.052						
ETHANE	C_2H_6	30.068						
PROPYLENE	C_3H_6	42.078						
PROPANE	C_3H_8	44.094						
HYDROGEN SULFIDE	H_2S	34.076						
CARBONYL SULFIDE	COS	60.075						
CARBON DISULFIDE	CS_2	76.13						
SULFUR DIOXIDE	SO_2	64.06						
NITROUS OXIDE	NO	30.008						
AMMONIA	NH_3	17.031						
HYDROGEN CYANIDE	HCN	27.026						
HYDROGEN CHLORIDE	HCl	36.461						
NAPHTHA	-							
TAR & OIL	-							
ASH	-							
OTHER SOLIDS								
SUBTOTAL, DRY								
WATER	H_2O	18.016	19.1					
TOTAL, WET			19.1	344				
GAS MOLECULAR WEIGHT								
TEMPERATURE, °F			100					
PRESSURE, PSIA			200					
SCFH								
GPM			1.0					

Table 4.12

Methanation Effluent Cooling Train - Texaco

ITEM NUMBER	NAME	DUTY 10^6 BTU/HR	EFF. ΔT °F	U	AREA FT^2
E-1	HP STEAM SUPERHEATER #2	46.4	246.8	60	3,135
E-2	HP STEAM GENERATOR #2	87.0	313	120	2,315
E-3	EXPANDER FEED GAS PREHEAT EXCHANGER	123.9	230	60	8,970
E-4	RECYCLE GAS PREHEAT EXCHANGER #1	50.0	197	60	4,220
E-5	HP STEAM GENERATOR #3	144.9	153	120	7,885
E-6	HP BFW PREHEATER	93.3	201	80	5,790
E-7	RECYCLE GAS PREHEAT EXCHANGER #2 A,B,C	395.0	100.3	60	21,880 (3 Shell)
EA-1	AIR COOLER	265.0	109	100	24,200
E-8	RECYCLE TRIM COOLER	100	30	80	375,000
E-9	150 PSIA STEAM GENERATOR	20	99	120	1,683
		2.6	96	80	3,375
EA-2	TRIM METHANA-TOR EFFLUENT AIR COOLER	23.4	130	100	1,802
E-10	TRIM COOLER	11.0	30	80	4,583

Table 4.13

Energy Balance, 10^6 BTU Per Hour
Texaco Process, System No.
Integrated SNG Module Definition

	PHASE	HHV	LATENT	SENSIBLE	TOTAL
STREAM Coal	Solid	4,575	14	0	4,575.0
Air to Air Sep'n	Gas		14	0	14
Raw Water	Liquid			0	0
SUBTOTAL IN					4,589.0
STREAM NET SNG	Gas	2,549.5		0	2,549.5
Ash	Solid	30.4		8.0	38.4
Acid Gas	Gas		3.2	1.5	4.7
Sulfur	Solid	61.5		0	61.5
Air Sep'n Vent	Gas		8.1	9.6	17.7
Flue Gases	Gas				
SUBTOTAL OUT					2,671.8
[illegible]					1,917.2
[illegible] WORK					
[illegible]					132.9
[illegible]					0
[illegible]			-712.5	-1,164.8	-1,877.3
[illegible]					0
[illegible]					-42.4
[illegible]					-1,786.5
[illegible]					1,786.5

[illegible] X 100% = 7.3%

[illegible]; Liquid Water

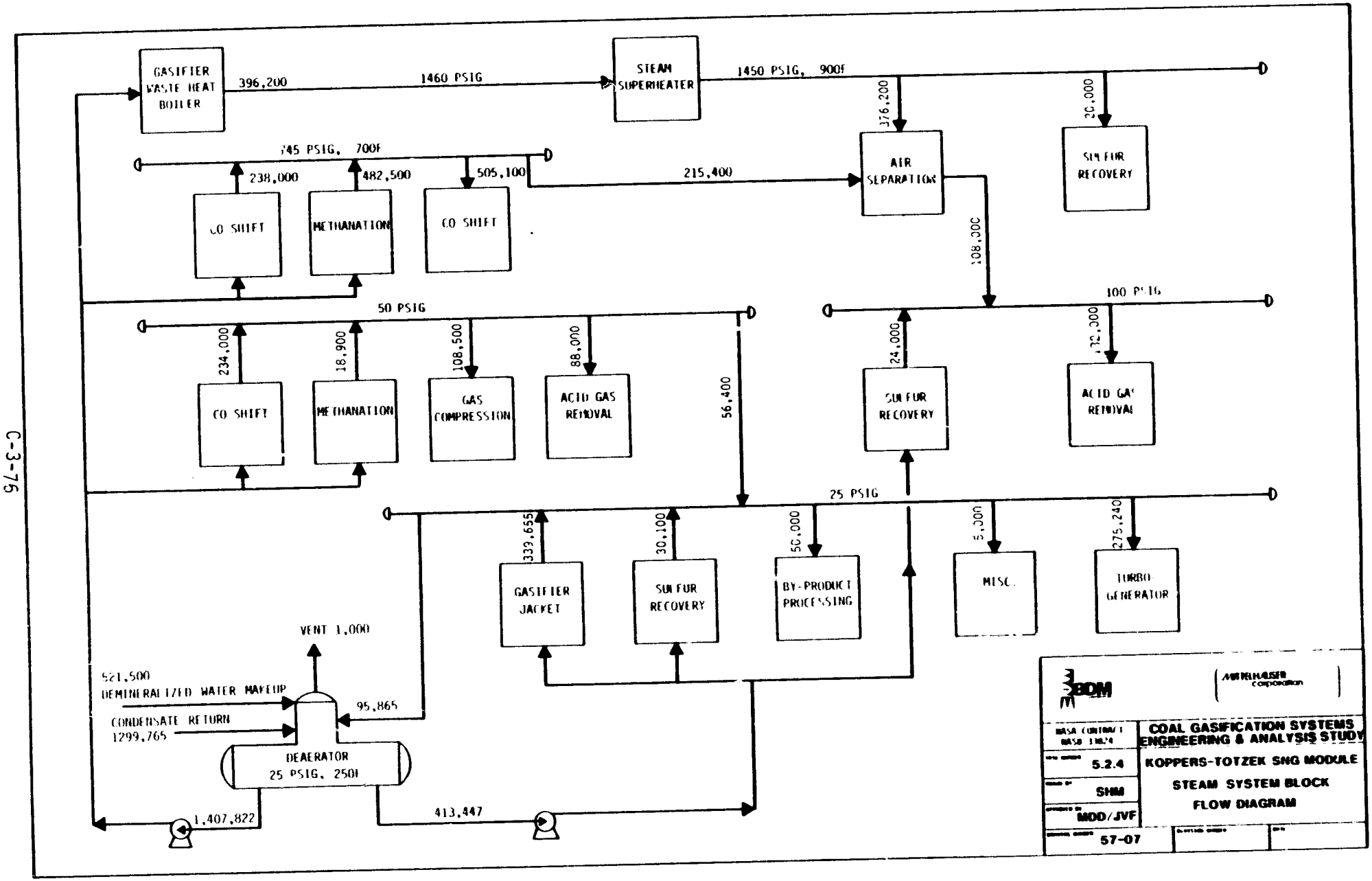
This balance represents one (1) Texaco MBG/SNG module

Table 4.14

Operating Requirements for Expected Operations
Texaco Process
Integrated SNG/Module Definition

	BASIS	UNITS
Raw Materials		1,825,000 TPY
Coal	TYP @ 100% operation	1.2 x 10^9 GALLONS/YEAR
Raw Water	GPY @ 100% operation	
Catalyst and Chemical Makeup		$1,572,000/YEAR
Makeup	100% Operation	$5,703,930
Initial Charge		
Operating Requirements		
Labor		$ 29,120
Supervisors	mh/Yr	195,864
Operators	mh/Yr	
Supplies	Factored as 15% of Operating Labor Costs	
Maintenance Requirements		
Labor	Factored as 1.6% of Total Depreciable Direct Investment	
Supplies	Factored as 2.4% of Total Depreciable Direct Investment	

The above operating requirements represent one (1) Texaco MBG/SNG module.


GASIFIER WASTE HEAT BOILER
396,200
1460 PSIG
STEAM SUPERHEATER
1450 PSIG, 900F
376,200
20,000
745 PSIG, 700F
238,000
482,500
505,100
215,400
AIR SEPARATION
SULFUR RECOVERY
CO SHIFT
METHANATION
CO SHIFT
108,300
100 PSIG
50 PSIG
234,000
18,900
108,500
88,000
24,000
CO SHIFT
METHANATION
GAS COMPRESSION
ACID GAS REMOVAL
56,400
SULFUR RECOVERY
ACID GAS REMOVAL
25 PSIG
339,655
30,100
50,000
5,300
275,240
GASIFIER JACKET
SULFUR RECOVERY
BY-PRODUCT PROCESSING
MISC.
TURBO GENERATOR
VENT 1,000
521,500
DEMINERALIZED WATER MAKEUP
95,865
CONDENSATE RETURN
1299,765
DEAERATOR
25 PSIG, 250F
1,407,822
413,447
BDM
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
KOPPERS-TOTZEK SNG MODULE
STEAM SYSTEM BLOCK FLOW DIAGRAM
SHM
MDD/JVF
57-07

for the facility, showing the impacts attributable to the MBG upgrading plant. Table 4.15 summarizes the impacts on the other systems in the MBG facility.

Koppers-Totzek/Texaco MBG Upgrading Module

Design No. 2 consists of one K-T MBG upgrading plant module, plus one Texaco MBG upgrading plant module, serving an MBG facility consisting of one K-T and three Texaco MBG modules. Drawing 57-08 is a diagram of system 15, Steam Generation/Distribution, for the facility, showing the impacts attributable to the MBG upgrading plant. Table 4.16 summarizes the impacts on the other systems in the MBG facility.

Table 4.15

Koppers-Totzek SNG Utility Summary

ITEM NO.	DISTRIBUTION	EQUIP. DUTY	COOLING WATER		ELECTRICITY		BOILER FEEDWATER		COND	STEAM DISTRIBUTION 1000 POUND/HR					
		10^6 BTU/HR	GPM		KW		GPM		GPM	HP PSIA		MP PSIA		LP PSIA	
			NORM	DESIGN	NORM	DESIGN	HP	LP		PROD	CONS	PROD	CONS	PROD	CONS
1	COAL PREPARATION														
2	GASIFICATION					9,250	792	679		396.0				339.7	
3	GAS COOLING														
4	ACID GAS REMOVAL			3,400		2,105		37							220
5	SULFUR RECOVERY			926		208	40	108			20.0		215.4	54.1	
6	AIR SEPARATION			28,066		3,040	752	438			376.0		215.4		
7	GAS COMPRESSION			400		78,790					20.0				108.5
8	PROCESS SOLIDS TREAT.														
11	COAL HANDLING					1,150									
12	SOLIDS DISPOSAL														
13	BY PRODUCT PROCESSING					60		100							50.0
14	PLANT POWER SYSTEM					(4,800)									275.3
15	STEAM GENERATION							191							95.9
16	WATER TREATMENT							10							5.0
17	COOLING WATER SYSTEM					1,830									
20A	CO SHIFT			1,900		375	476		1,022			238	505.1	234	
20B	METHANATION			3,533					233			482.5		18.9	
	GENERAL SUPPORT					5,006									

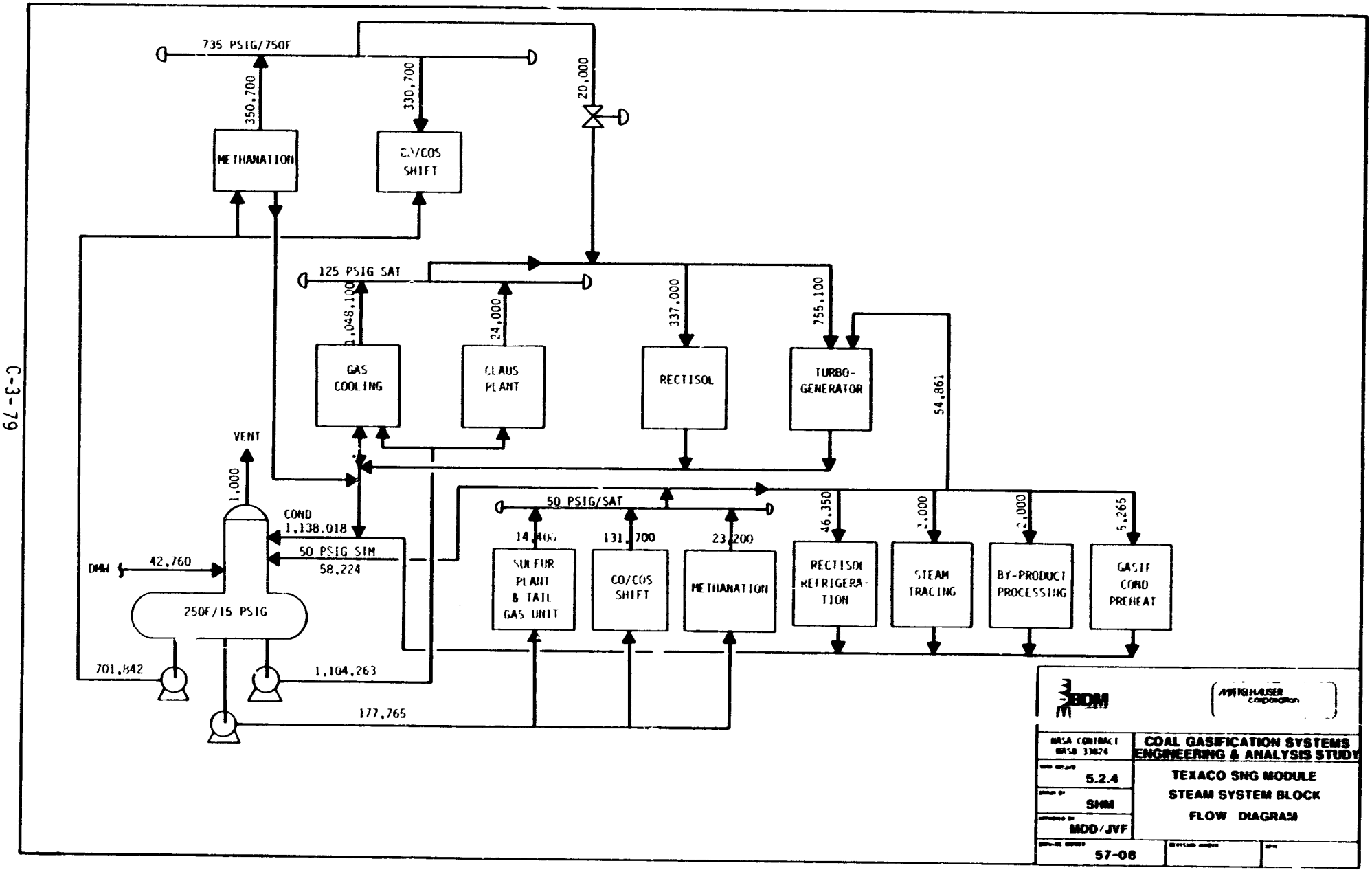

735 PSIG/750F
350,700
330,700
20,000
METHANATION
CO/COS SHIFT
125 PSIG SAT
1,048,100
24,000
337,000
755,100
GAS COOLING
CLAUS PLANT
RECTISOL
TURBO-GENERATOR
54,861
VENT
1,000
COND 1,138,018
50 PSIG/SAT
14,400
131,700
23,200
46,350
2,000
2,000
5,265
DMW
42,760
50 PSIG SIM 58,224
SULFUR PLANT & TAIL GAS UNIT
CO/COS SHIFT
METHANATION
RECTISOL REFRIGERATION
STEAM TRACING
BY-PRODUCT PROCESSING
GASIF COND PREHEAT
250F/15 PSIG
701,842
1,104,263
177,765
BDM
MITTELHAUSER corporation
NASA CONTRACT NASW 3307A
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
TEXACO SNG MODULE
STEAM SYSTEM BLOCK
FLOW DIAGRAM
SNM
MDD/JVF
57-08

Table 4.16

Texaco-SNG Utility Summary

ITEM NO.	DISTRIBUTION	EQUIP. DUTY	COOLING WATER		ELECTRICITY		BOILER FEEDWATER		COND	STEAM DISTRIBUTION 1000 POUND/HR					
										HP PSIA		MP PSIA		LP PSIA	
		10^6 BTU/HR	GPM		KW		GPM		GPM	PROD	CONS	PROD	CONS	PROD	CONS
			NORM	DESIGN	NORM	DESIGN	HP	LP							
1	COAL PREPARATION					3,000									
2	GASIFICATION			3,430		200	1,778	318						1,048.1	
3															
4	ACID GAS REMOVAL			31,000		2,173							46.3		33.7
5	SULFUR RECOVERY			800			48	29				24		144	
6	AIR SEPARATION			2,200		68,000									
8	PROCESS SOLIDS TREAT.														
11	COAL HANDLING					1,890									
12	SOLIDS DISPOSAL														
13	BYPRODUCT PROCESSING					60									2
14	PLANT POWER SYSTEM			39,400		(38,600)							755.1		54.9
15	STEAM GENERATION														58.2
16	WATER TREATMENT														
17	COOLING WATER SYSTEM					4,200									
20A	CO SHIFT			4,320		375	347		690			173.7	504.4	131.7	
20B	METHANATION			8,567		20,368	707		239			330.7		23.2	
	GENERAL SUPPORT					18,477									6.4

SUPPLEMENT A TO APPENDIX C-3

KOPPERS-TOTZEK
SNG MODULE
EQUIPMENT LIST

EQUIPMENT LIST
KOPPERS-TOTZEK
SNG-MODULES

UNIT: CO SHIFT REV. 1 DATE: 10/6/80

EXCHANGERS

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
90-E-1	HP STEAM SUPER HEATER	0	1	AREA = 2096 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 800 PSIA SHELL - 750 PSIA MATERIAL - CS/MOLY
90-E-2	HP STEAM GENERATOR	0	1	AREA = 10,208 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 750 PSIA SHELL - 800 PSIA MATERIAL - CS/MOLY
90-E-3	BFW PREHEAT EXCHANGER	0	1	AREA = 7893 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 750 PSIA SHELL - 800 PSIA MATERIAL - CS/MOLY
90-E-4	BFW PREHEAT EXCHANGER	0	1	AREA = 9900 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 750 PSIA SHELL - 650 PSIA MATERIAL - CS/CS
90-EA-1	SHIFT EFFLUENT COOLER	0	1	AREA = 24,500 FT^2 TYPE - AIR COOLER DES PRESS: TUBE - 650 PSIA MATERIAL - CS TUBE LENGTH = 20'
90-E-5	SHIFT FEED/EFFLUENT EXCHANGER	0	1	AREA = 1890 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 150 PSIA SHELL - 650 PSIA MATERIAL - CS/CS
90-E-6	65 PSIA STEAM GENERATOR	0	1	AREA = 18,560 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 650 PSIA SHELL - 650 PSIA MATERIAL - CS/CS
90-E-7	SHIFT EFFLUENT TRIM COOLER	0	2	AREA = 31,250 FT^2 TYPE - AIR COOLER DES PRESS: TUBE - 650 PSIA MATERIAL - CS/ADMIRALTY

EQUIPMENT LIST
KOPPERS-TOTZEK
SNG-MODULES
REV. 1

UNIT: CO SHIFT

DATE: 10/6/80

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
90-E-8	HYDROLYSIS EFFLUENT TRIM COOLER	0		AREA = 1200 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 650 PSIA SHELL -650 PSIA MATERIAL - CS/CS
90-EX-2	HYDROLYSIS EFFLUENT COOLER	0		AREA = 2020 FT^2 TYPE - AIR COOLER DES PRESS: TUBE - 650 PSIA TUBE LENGTH - 20' MATERIAL - CS
90-E-9	HYDROLYSIS FEED/EFFLUENT EXCHANGER	0	1	AREA = 4375 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 150 PSIA SHELL -650 PSIA MATERIAL - CS/ADMIRALTY
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -

EQUIPMENT LIST
KOPPERS-TOTZEK
SNG-MODULES
REV. 1

UNIT: METHANATION

DATE: 10/6/80

COMPRESSORS				
ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
90-EX-1	METHANATION FEED GAS EXPANDER	0	1	GAS HORSEPOWER - 13,118 TYPE - CENTRIFUGAL MATERIAL = 5 CR STEEL ΔP - 308 PSI
90-M-1	RECYCLE COMPRESSOR ASSIST MOTOR DRIVER	0	1	BRAKE Hp - 7887 TYPE - ELECTRIC MOTOR
90-C-1	METHANATION RECYCLE COMPRESSOR	0	1	GAS HORSEPOWER - 21,005 TYPE - CENTRIFUGAL MATERIAL - 5 CR STEEL ΔP - 60 PSI
90-M-2	RECYCLE COMPRESSOR MOTOR DRIVER (START-UP)	0	1	BRAKE Hp - 21,005 TYPE - ELECTRIC MOTOR
90-C-2	METHANE PRODUCT COMPRESSOR			GAS HORSEPOWER = 6800 TYPE - CENTRIFUGAL MATERIAL - CAST STEEL ΔP = 800 PSI
90-ST-1	PRODUCT COMPRESSOR STEAM TURBINE DRIVER			BRAKE Hp - 6800 TYPE - CONDENSING TURBINE MATERIAL - 1500 PSI

EQUIPMENT LIST
KOPPERS-TOTZEK
SNG-MODULES
REV. 1

UNIT: CO SHIFT

DATE: 10/6/80

VESSELS

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
90-RX-1	CO SHIFT REACTOR	0	1	DIAMETER = 18'-0" HEIGHT = 18'-0" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 750 PSIA
90-RX-2	COS HYDROLYSIS REACTOR	0	1	DIAMETER = 6'-6" HEIGHT = 6'-6" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 750 PSIA
90-V-1	SHIFT EFFLUENT CONDEN-SATE KNOCK-OUT DRUM	0	1	DIAMETER = 7'-3" HEIGHT = 18'-9" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 650 PSIA
90-V-2	HYDROLYZER EFFLUENT CONDENSATE KNOCK-OUT DRUM	0	1	DIAMETER = 2'-6" HEIGHT = 6'-9" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 690 PSIA
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =

EQUIPMENT LIST
KOPPERS-TOTZEK
SNG-MODULES

UNIT: METHANATION REV. 1 DATE: 10/6/80

VESSELS

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
92-RX-1	ZINC OXIDE GUARD BED	1	1	DIAMETER = 10'-8" HEIGHT = 26'-8" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 325 PSIA
92-RX-2	METHANATION REACTOR	0	1	DIAMETER = 21'-6" HEIGHT = 21'-6" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 350 PSIA
92-RX-3	TRIM METHANATION REACTOR	0	1	DIAMETER = 7'-8" HEIGHT = 7'-8" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 350 PSIA
92-V-1	RECYCLE COMPRESSOR SUCTION DRUM	0	1	DIAMETER = 16'-3" HEIGHT = 40'-8" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 350 PSIA
92-V-2	METHANE COMPRESSOR SUCTION DRUM	0	1	DIAMETER = 4'-0" HEIGHT = 10'-0" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 350 PSIA
92-V-3	METHANE COMPRESSOR INTERSTAGE KNOCK-OUT DRUM	0	1	DIAMETER = 2'-6" HEIGHT = 6'-3" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 750 PSIA
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =

EQUIPMENT LIST
KOPPERS-TOTZEK
SNG-MODULES
REV. 1

UNIT: METHANATION

DATE: 10/6/80

EXCHANGERS

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
92-E-2	METHANATOR FEED/ EFFLUENT EXCHANGER	0	1	AREA = 10,480 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 700 PSIA SHELL -350 PSIA MATERIAL - CS/MOLY
92-E-1	HP STEAM GENERATOR	0	1	AREA = 4660 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 750 PSIA SHELL - 350 PSIA MATERIAL - MOLY/MOLY
92-E-4	HP BFW PREHEAT EXCHANGER	0	1	AREA = 3975 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 750 PSIA SHELL - 350 PSIA MATERIAL - CS/MOLY
92-E-3	565 PSIA STEAM GENERATOR	0	1	AREA = 6830 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 750 PSIA SHELL - 350 PSIA MATERIAL - CS/CS
92-E-5 A-D	RECYCLE/EFFLUENT EXCHANGER	0	4	AREA = 32,680 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 350 PSIA SHELL - 350 PSIA MATERIAL - CS/CS
92-EA-1	EFFLUENT AIR COOLER	0	1 (15 BAYS)	AREA = 15,550 FT^2 DESIGN PRESS = 350 MATERIAL - CS
92-E-6 A,B,C	EFFLUENT TRIM COOLER	0	3	AREA = 11,250 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 150 PSIA SHELL - 300 PSIA MATERIAL - CS/ADMIRALTY
92-E-7	LP PSIA STEAM GENERATOR	0	1	AREA = 800 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 350 PSIA SHELL - 200 PSIA MATERIAL - CS/CS

EQUIPMENT LIST
KOPPERS-TOTZEK
SNG-MODULES

UNIT: METHANATOR REV. 1 DATE: 10/8/80

EXCHANGERS				
ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
92-EA-2	TRIM METHANATOR AIR COOLER	0	1	AREA = 1330 FT^2 DESIGN PRESS - 350 MATERIAL - CS
92-E-8	TRIM METHANATOR TRIM COOLER	0	1	AREA = 2750 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 150 SHELL -350 MATERIAL - CS/ADMIRALTY
92-E-9	METHANE COMPRESSOR INTERCOOLER	0	1	AREA = 1413 FT^2 TYPE - U-TUBE DES PRESS: TUBE - SHELL - MATERIAL - CS/ADMIRALTY
92-E-10	METHANE PRODUCT COOLER	0	1	AREA = 1953 FT^2 TYPE - U-TUBE DES PRESS: TUBE - SHELL - MATERIAL - CS/ADMIRALTY
92-E-11	STEAM TURBINE SURFACE CONDENSOR	0	1	PART OF COMPRESSOR COST
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -

SUPPLEMENT B TO APPENDIX C-3

TEXACO - SNG MODULE
EQUIPMENT LIST

EQUIPMENT LIST
TEXACO
SNG-MODULES
REV. 1

UNIT: CO SHIFT DATE: 10/6/80

EXCHANGERS

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
90-E-1	HP STEAM GENERATOR	0	1	AREA = 1200 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 750 PSIA SHELL - 690 PSIA MATERIAL - CS/MOLY
90-E-4	FEED CO SHIFT EFFLUENT EXCHANGER	0	1	AREA = 8072 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 690 PSIA SHELL -690 PSIA MATERIAL - CS/MOLY
90-E-3	BFW PREHEAT EXCHANGER	0	1	AREA = 2440 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 750 PSIA SHELL - 690 PSIA MATERIAL - CS/MOLY
90-E-5	65 PSIA STEAM GENERATOR	0	1	AREA = 11,655 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 135 PSIA SHELL - 650 PSIA MATERIAL - CS/CSX
90-EA-1	SHIFT EFFLUENT COOLER	0	1 (18 BAYS)	AREA = 21,925 FT^2 TYPE - AIR COOLER DES PRESS: TUBE - 650 PSIA MATERIAL - CS TUBE LENGTH = 20'
90-E-6	SHIFT EFFLUENT TRIM COOLER	0	2	AREA = 37,500FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 650 PSIA SHELL - 150 PSIA MATERIAL - CS/ADMIRALTY
90-E-7	HYDROLYSIS REACTOR FEED/ EFFLUENT EXCHANGER	0	1	AREA = 1740 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 650 PSIA SHELL - 650 PSIA MATERIAL - CS/CS
90-EA-2	HP STEAM GENERATOR	0	1	AREA = 6850 FT^2 TYPE - DES PRESS TUBE - 50 PSIA SHELL - 590 PSIA MATERIAL - CS/MOLY

EQUIPMENT LIST
TEXACO
SNG-MODULES

UNIT: CO SHIFT   REV. 1   DATE: 10/6/80

EXCHANGERS				
ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
90-E-8	HYDROYLSIS EFFLUENT TRIM COOLER	0	3	AREA = 2000 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 150 PSIA SHELL - 650 PSIA MATERIAL - CS/ADMIRALTY
90-EA-2	HYDROLYSIS EFFLUENT COOLER		1 (2 BAYS)	AREA = 2545 FT^2 TYPE - AIR COOLER DES PRESS: TUBE - 650 PSIA TUBE LENGTH - 20' MATERIAL - CS
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -

EQUIPMENT LIST
TEXACO
SNG-MODULES
REV. 1

UNIT: CO SHIFT

DATE: 10/6/80

VESSELS

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
90-RX-1	CO SHIFT REACTOR	0	1	DIAMETER = 18'-0" HEIGHT = 18'-0" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS =
90-RX-2	COS HYDROLYSIS REACTOR	0	1	DIAMETER = 6'-6" HEIGHT = 6'-6" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS =
90-V-1	SHIFT CONDENSATE KNOCK-OUT DRUM	0	1	DIAMETER = 7'-3" HEIGHT = 18'-9" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 650 PSIA
90-V-2	HYDROLYSIS CONDENSATE KNOCK-OUT DRUM	0	1	DIAMETER = 2'-6" HEIGHT = 6'-9" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 690 PSIA
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =

EQUIPMENT LIST
TEXACO
SNG-MODULES
REV. 1

UNIT: METHANATION DATE: 10/6/80

EXCHANGERS

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
92-E-1	HP STEAM SUPERHEATER	0	1	AREA = 3135 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 700 SHELL -350 MATERIAL - CS/MOLY
92-E-2	HP STEAM GENERATOR	0	1	AREA = 2315 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 750 SHELL - 350 MATERIAL - MOLY/MOLY
92-E-3 A,B	EXPANDER FEED GAS PREHEAT EXCHANGER	0	2	AREA = 8970 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 650 SHELL - 350 MATERIAL - CS/MOLY
92-E-5	HP STEAM GENERATOR	0	1	AREA = 7885 FT^2 TYPE - KETTLE REBOILER DES PRESS: TUBE - 750 SHELL - 350 MATERIAL - CS/CS
92-E-4	RECYCLE/GAS PREHEAT EXCHANGER	0	2	AREA = 4220 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 350 SHELL - 350 MATERIAL - CS/CS
92-EA-1	EFFLUENT AIR COOLER	0	1 (9 BAYS)	AREA = 24,200 FT^2 DESIGN PRESS = 350 MATERIAL - CS TUBE LENGTH - 20'
92-E-6	BFW PREHEATER EXCHANGER	0	1	AREA = 5790 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 750 SHELL - 300 MATERIAL - CS/CS
92-E-7	RECYCLE GAS PREHEAT EXCHANGER	0	3	AREA = 21,880 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 350 SHELL - 350 MATERIAL - CS/CS

EQUIPMENT LIST
TEXACO
SNG-MODULES

EXCHANGERS				
ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
92-E-2	TRIM METHANATOR AIR COOLER	0	1	AREA = 1800 FT^2 DESIGN PRESS - 350 MATERIAL - CS
92-E-10	TRIM METHANATOR TRIM COOLER	0	1	AREA = 4580 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 150 SHELL -350 MATERIAL - CS/ADMIRALTY
92-E-11	METHANE COMPRESSOR INTERCOOLER	0	1	AREA = 1413 FT^2 TYPE - U-TUBE DES PRESS: TUBE - SHELL - MATERIAL - CS/ADMIRALTY
92-E-12	METHANE PRODUCT COOLER	0	1	AREA = 1953 FT^2 TYPE - U-TUBE DES PRESS: TUBE - SHELL - MATERIAL - CS/ADMIRALTY
92-E-11	STEAM TURBINE SURFACE CONDENSOR	0	1	PART OF COMPRESSOR COST
92-E-8	RECYCLE TRIM COOLER	0	1	AREA = 41,700 FT^2 TYPE - U-TUBE DES PRESS: TUBE - 350 PSIA SHELL -150 PSIA MATERIAL - CS/ADMIRALITY
92-E-9	150 PSIA STEAM GENERATOR	0	1	AREA = 2020 FT^2 TYPE - FIXED TUBE SHEET DES PRESS: TUBE - 350 PSIA SHELL - 200 PSIA MATERIAL - CS/CS
				AREA = TYPE - DES PRESS: TUBE - SHELL - MATERIAL -

EQUIPMENT LIST
TEXACO
SNG-MODULES

ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
92-EX-1	METHANATION FEED GAS EXPANDER	0	1	GAS HORSEPOWER - 13,118 TYPE - CENTRIFUGAL MATERIAL = 5 CR STEEL ΔP - 308 PSI
92-M-2	EXPANDER MOTOR DRIVER	0	1	BRAKE Hp - 4400 TYPE - ELECTRIC MOTOR
92-C-1	METHANATION RECYCLE COMPRESSOR	0	1	GAS HORSEPOWER - 17,402 TYPE - CENTRIFUGAL MATERIAL - 5 CR STEEL ΔP - 60 PSI
92-M-1	RECYCLE COMPRESSOR MOTOR DRIVER	0	1	BRAKE Hp - 18,000 TYPE - ELECTRIC MOTOR
92-C-2	METHANE PRODUCT COMPRESSOR			GAS HORSEPOWER = 6800 TYPE - CENTRIFUGAL MATERIAL - CAST STEEL ΔP = 800 PSI
92-ST-1	PRODUCT COMPRESSOR STEAM TURBINE DRIVER			BRAKE Hp - 7000 TYPE - CONDENSING TURBINE MATERIAL - 1500 PSI

EQUIPMENT LIST
TEXACO
SNG-MODULES

VESSELS				
ITEM NUMBER	SERVICE	QUANTITY		DESCRIPTION
		SPARE	OPERATION	
92-RX-1	ZINC OXIDE GUARD BED	1	1	DIAMETER = 10'-8" HEIGHT = 26'-8" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 325 PSIA
92-RX-2	METHANATION REACTOR	0	1	DIAMETER = 21'-6" HEIGHT = 21'-6" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 350 PSIA
92-RX-3	TRIM METHANATION REACTOR	0	1	DIAMETER = 7'-8" HEIGHT = 7'-8" TYPE - VERTICAL MATERIAL - 304 SS CLAD DESIGN PRESS = 350 PSIA
92-V-1	RECYCLE COMPRESSOR SUCTION KNOCK-OUT DRUM	0	1	DIAMETER = 16'-3" HEIGHT = 40'-8" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 350 PSIA
92-V-2	METHANE COMPRESSOR SUCTION KNOCK-OUT DRUM	0	1	DIAMETER = 4'-0" HEIGHT = 10'-0" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 350 PSIA
92-V-3	METHANE COMPRESSOR INTERSTAGE KNOCK-OUT DRUM	0	1	DIAMETER = 2'-6" HEIGHT = 6'-3" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 750 PSIA
92-V-4	METHANE COMPRESSOR AFTER COOLER	0	1	DIAMETER = 2'-6" HEIGHT = 6'-3" TYPE - HORIZONTAL MATERIAL - CS DESIGN PRESS = 1100 PSIA
				DIAMETER = HEIGHT = TYPE - MATERIAL - DESIGN PRESS =

REFERENCES

1. "Koppers-Totzek Process Reference Facility Design," Volumes II-III, by The BDM Corporation, Contract NAS8-33824 dated July 21, 1980.

2. "Texaco Process Reference Facility Design", Volumes I-II, by The BDM Corporation, Contract NAS8-33824, dated September 15, 1980.

3. "Cost Estimation and Economic Evaluation Methodology, Task 5.4.2.3", by The BDM Corporation, Contract NAS8-33824, dated September 19, 1980.

4. "Coal Gasification Systems Engineering and Analysis Task 5.1.4, Alternate Products Analysis, K-T and Texaco Cases," by The BDM Corporation, Contract NAS8-33824, dated July 24, 1980.

5. "Coal Gasification Systems Engineering and Analysis, Task 5.7.1, Alternate Products Analysis, Lurgi Case" by The BDM Corporation, Contract NAS8-33824, dated November 5, 1980.

6. "Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant", by the Tennessee Valley Authority, March 1980.

7. "Factored Estimates for Western Coal Commercial Concepts-Interim Report, Appendix A - Coal Gasification Commercial Concepts Gas Cost Guidelines", C.F. Braun & Co., Report No FE-2240-5, October 1976.

8. Briefing Slides, The BDM Corporation, presented at NASA Briefing of May 29, 1980, Contract NAS8-33824.

9. "Study Plan for a Coal Gasification Systems Engineering and Analysis Study," by The BDM Corporation, Contract NAS8-33824, dated April 1980.

10. "Conceptual Design for Advanced Coal Liquefaction Commercial Plant - Final Report, Additional Studies," by Fluor Engineers Constructors, Incorporated, Report No. FE-2251-52, June 1978.

11. "Factored Estimates for Eastern Coal Commercial Concepts, Interim Report," C.F. Braun & Co., Report No. FE-2240-31, September 1978.

12. "Carbon Monoxide Shift Catalysts and Carbonyl Sulfide Hydrolysis Catalysts," by I. Dybkgaer, presented at the TVA Ammonia from Coal Symposium, May 8-10, 1979.

13. "Phase I: The Pipeline Gas Demonstration Plant, Design and Evaluation of Commercial Plant, Volume 2- Process and Project Engineering Design," by Continental Oil Company, Report No. 2542-10 Vol. 2, 1978.